THE DUAL DIGESTION OF SEWAGE SLUDGE USING AIR AND PURE OXYGEN

BY

ANDREW J PITT BSc(Hons)
University of Sheffield

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Department of Civil Engineering
University of Cape Town

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DECLARATION BY CANDIDATE

I, Andrew James Pitt, hereby declare that this thesis is my own work and that it has not been submitted for a degree at another University.

Signed by candidate

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In Memory of my Father

James John Pitt

1921-1991

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ABSTRACT

Dual Digestion is a two stage system that combines autothermal thermophilic aerobic pre-treatment with conventional anaerobic digestion. The practicability of the system using pure oxygen is well proven. Disadvantages are the high cost of the pure oxygen and the absence of a detailed evaluation of anaerobic digester performance. This report discusses the results of a full-scale investigation into the dual digestion system ($184m^3$ aerobic reactor and $1800m^3$ anaerobic digester), carried out in two phases: In the first using air alone for oxygenating the aerobic reactor and in the second using a combination of air and pure oxygen. During both phases the performance of the anaerobic digester was also monitored, but in greater detail in the second phase as far as the final sludge product is concerned.

In phase I, with air, it was possible to maintain thermophilic temperatures in the aerobic reactor throughout the year. However the required retention times were relatively long (3-6 days) in comparison with the pure oxygen reactor (~1 day) due to the high vapour heat losses. At long retention times, the volatile solids (VS) destruction was appreciable (~25%) and the reactor tended towards an autothermal thermophilic digester. Foaming, although unpredictable in its occurrence, significantly improved aerobic reactor performance by doubling the oxygen transfer efficiency. From liquid and gas mass and heat balances it was found that the specific biological heat yield and respiration quotient were approximately constant at 12.8 MJ/kg(O_2) and 0.70 mol(CO_2)/mol(O_2) respectively over a wide range of operating conditions and consistent relationships between VS removal, heat generation, and oxygen utilisation could be established. Based on information collected, it was concluded that increased treatment capacity and greater temperature control of the aerobic reactor could be provided by supplementing air oxygenation with pure oxygen.

In phase II, using a combination of air and pure oxygen, much higher loading rates on the aerobic reactor were possible. Thermophilic temperatures could be maintained at short retention times (1-2 days). Unfortunately no foaming occurred during this period. Consequently the benefit of improved oxygen transfer efficiency of the air oxygenation system, produced by the foam, could not be exploited. Liquid and gas mass and heat balances confirmed the specific heat yield and respiration quotient values and the relationship between oxygen utilisation, VS destruction and biological heating. During phase II, the anaerobic digester operated at a retention time of ~10 days. The sensible heat content of the hot sludge from the aerobic reactor was sufficient to force the digester into the thermophilic temperature range. The stability of the anaerobic process and final

sludge product at this short retention time was monitored with %VS removal and residual specific oxygen utilisation rate tests and found to be similar to that of conventional mesophilic anaerobic digestion at 20 days retention time. Dewaterability as reflected by the specific resistance to filtration (SRF) was found to be poor, but not much worse than for conventional mesophilic digestion.

Sufficient information was obtained during phases I and II to allow a mathematical model to be compiled, which could reasonably reliably simulate all the main operating parameters of the dual digestion system. The model provided a means for assessing different system configurations with mesophilic or thermophilic digestion, with and without heat exchange or gas engine external heat sources, allowing technical and economical (capital and operating) feasibility to be evaluated and compared with that for conventional digestion.

From both the experimental and modelled results, all the claimed benefits of the dual digestion system were verified with the exception of the claim that aerobic reactor heat pre-treatment of the sludge allows the anaerobic digester to operate at short retention times (~10 days). However, the digester can be operated at 10 days retention provided its temperature is in the thermophilic range, in which case a sufficiently stable sludge is produced; at mesophilic temperatures, a retention time of 15 days or longer is required to produce a sludge of equivalent stability to that from conventional mesophilic digestion. Consequently it is not the stability of the anaerobic process per se that governs the minimum retention time but the quality required for the final sludge product. The aerobic reactor is an appropriate pre-treatment stage for the thermophilic digester because it provides the necessary temperature and pH buffering to allow stable operation in the thermophilic range.

It is concluded that where application of conventional anaerobic digestion is contemplated, whether for new installations or for upgrading existing plants, the dual digestion system should be seriously considered as a possible option. It competes favourably both technically and economically with conventional mesophilic digestion and produces a superior sludge product which can be beneficially utilised in agriculture.

Andrew James Pitt
Department of Civil Engineering
University of Cape Town
Private Bag Rondebosch
Cape Town 7700
South Africa

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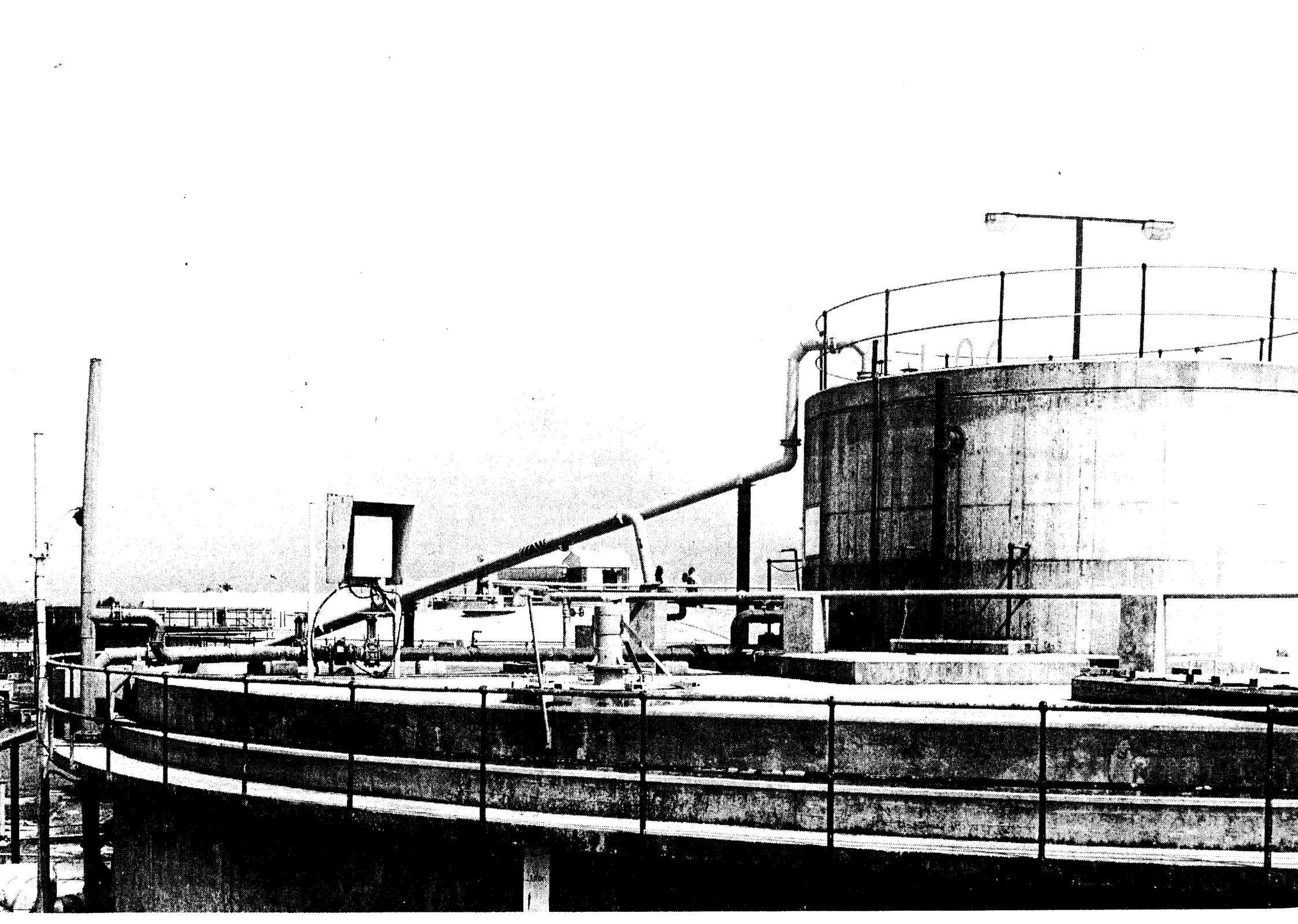

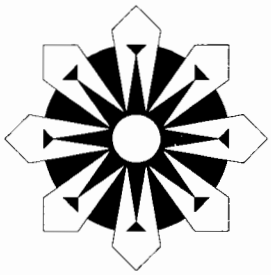

SYNOPSIS

INTRODUCTION

Sludge pasteurisation and increased anaerobic digester treatment capacity are important objectives in municipal sewage sludge treatment. At Milnerton, Cape Town, a pure oxygen dual digestion plant, comprising a first stage autothermal thermophilic (60°C) aerobic reactor followed by a second stage mesophilic (37°C) anaerobic digester, demonstrated that this system was capable of meeting both these objectives (Messenger *et al*, 1993). In addition to providing pasteurisation in a simple to operate process, the autothermal heat treatment in the aerobic reactor conditioned the sludge allowing the anaerobic digester to be operated at shorter retention times, thereby increasing the treatment capacity of the anaerobic digester. However, disadvantages of the Milnerton dual digester were (1) the high cost of the pure oxygen for aerating the aerobic reactor (Laubscher *et al*, 1992) and (2) the quality of the final sludge produced (Messenger *et al*, 1992). For a number of reasons the anaerobic digester performance at Milnerton could not be satisfactorily evaluated. If it is possible both technically and economically to operate the reactor with air, and the anaerobic digester performance and product properly evaluated, then dual digestion may prove to be a simple and inexpensive option for improving sludge treatment in anaerobic digestion.

OBJECTIVES PHASE I: DUAL DIGESTION USING AIR ALONE

In 1989, the Cape Town City Council initiated a full scale research project to investigate the dual digestion system employing air rather than oxygen for aerating the aerobic reactor. While the principal objective was to demonstrate the practicability of the system as a means of increasing anaerobic digester treatment capacity, the investigation examined all the claimed benefits of the system viz:

- aerobic conditioning which accelerates the anaerobic digestion process
- sludge pasteurisation and pathogen inactivation without adversely effecting the anaerobic digestion process such as reducing biogas production and VS removal
- no external heating is required for the anaerobic digester

Apart from evaluating the above claims for the dual digestion system aerated with air, additional objectives of the investigation were to:

- define the aeration and heating requirements to achieve autoheating with air aeration and to establish the minimum practical aerobic and anaerobic retention times for the system;
- develop a mathematical method based on liquid and gas heat and mass balances for the system allowing prediction of aerobic reactor performance such as reactor temperature, VS removal, etc.
- evaluate practical operating problems of the system at full scale over an extended period to assess the reliability and dependability of the system.
- estimate capital, operation and maintenance costs for the system and compare these with conventional anaerobic digestion.

OPERATION AND MONITORING: PHASE I

The dual digester was constructed by building a 184m^3 liquid volume aerobic reactor inside an existing (nominal 2000m^3) anaerobic digester, leaving a liquid volume of 1800m^3 for anaerobic digestion. Details of the plant layout are given in Figure 1.

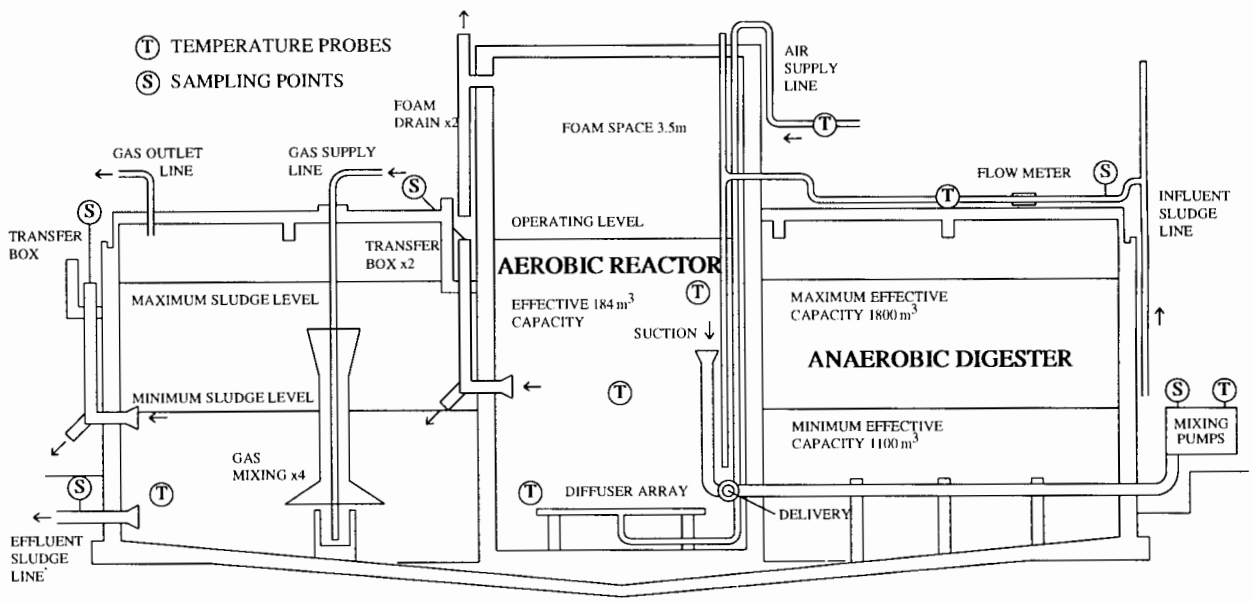

FIGURE 1 SCHEMATIC OF THE DUAL DIGESTION PLANT AT ATHLONE

The aerobic reactor was aerated by one 20kW liquid ring compressor, producing a maximum dry air flow rate of 760m^3 (STP)/h, feeding course bubble diffusers set in a ring at the bottom of the reactor. Mixing was augmented with a 10kW recirculation pump, discharging 1000m^3/h in a horizontal and tangential pattern into the base of the reactor. The anaerobic digester was mixed by recirculating biogas, pumped by a second identical liquid ring compressor. The capacity of the anaerobic digester was too large to allow short retention times to be tested during phase I and consequently this was done at laboratory scale. The aerobic reactor was fed with gravity thickened primary

sludge (~4%TS) in batches at 2 to 4 hourly intervals. Influent feed sludge displaced reactor sludge to the anaerobic digester, which in turn displaced sludge to a secondary digester. Feeding was not performed on a draw and fill basis as with most pasteurisation systems. It was recognised that the displacement type of feeding procedure would lead to short circuiting and poor pathogen inactivation performance, but in this prototype plant, this was not regarded as a serious problem.

All parameters to allow (liquid and gas) heat and mass balances to be made across the aerobic reactor were measured. This involved measuring many of the influent and effluent liquid and gas flow rates, constituents and temperatures. The materials considered were sludge mass dry solids (Total Solids, Volatile Solids and COD), water, oxygen, nitrogen and carbon dioxide. VS removal in the anaerobic digester was monitored. However, as there were no flow meters to measure the gas flow rate in the biogas recirculation line, it was not possible to conduct gas mass and heat balances on the digester.

RESULTS FROM THE INVESTIGATION: PHASE I

Overall System Performance

The dual digester was monitored daily for a period of 312 days during phase I, incorporating a full summer and winter season. During this time various operating parameters, such as the air and influent sludge flow rates, were changed to examine the influence of these parameters on the performance of the system. A summary of performance over the 312 day period is given in Tables 1-4.

TABLE 1 CHEMICAL CHARACTERISTICS PHASE I

Chemical Characteristics†		Feed Sludge	Aerobic Sludge	Anaerobic Sludge
Temperature	°C	**20** (12-27)	**49** (40-59)	**31** (24-36)
Total Solids	g/ℓ	**45** (24-88)	**35** (15-55)	**24** (6-70)
Volatile Solids	g/ℓ	**37** (20-74)	**28** (12-47)	**17** (3-46)
COD	g(O)/ℓ	**64** (21-99)	**43** (15-79)	**26** (7-72)
pH	-	**5.4** (4.8-6.1)	**7.4** (6.1-8.1)	**7.4** (7.0-7.9)
Ammonia	mg(N)/ℓ	**110** (49-240)	**370** (41-700)	**760** (570-990)
Bicarbonate Alkalinity	mg($CaCO_3$)/ℓ	**40** (0-460)	**820** (0-1460)	**2910** (2160-3460)

† Mean values are quoted with the ranges given in brackets.

The system proved simple to operate and no major mechanical problems were encountered. The aerobic reactor was easily started and reached thermophilic temperatures (~50°C) within 10 days. It was important to ensure that the feed sludge was adequately concentrated to prevent the reactor from becoming substrate limited. A

significant foam layer (~3m deep) was present in the reactor at various stages during the investigation, particularly when the temperature was in excess of 50°C. Although the foam was difficult to control and could only be managed by reducing the air flow rate, its presence proved beneficial to the reactor doubling the oxygen transfer efficiency (*OTE*) of the air oxygenation device.

The aerobic reactor was able to reach thermophilic temperatures throughout the year although longer retention times were required in the winter months. The average retention time for the period was 4.4 days with the result that volatile solid reduction across the reactor was about 25%, a factor which would impact on the quantity of biogas produced in the digester. Because of the relatively large operating capacity of the anaerobic digester (1800m^3), the average retention time in the anaerobic digester was long (42 days). With insufficient sensible heat being provided by the sludge from the aerobic reactor, digester temperatures were generally below the mesophilic optimum (37°C) i.e. in the range 26°C (winter) to 33°C (summer). Overall removal of VS and COD by the system was 46% and 59% respectively. The conditioning effects of aerobic pre-treatment were evident from significant increases in the sludge ammonium and bicarbonate alkalinity concentrations, and pH after aerobic treatment.

TABLE 2 SOLIDS REMOVAL: PHASE I

Solids Removals†		Aerobic	Anaerobic	Overall
Total Solids	%	23	32	48
Volatile Solids	%	25	40	46
COD	%	33	40	59

† Mean values are quoted.

In the laboratory scale study, to determine the minimum anaerobic digester retention time for process stability, a mesophilic (37°C) anaerobic digester fed with aerobic reactor sludge satisfactorily operated at an 8 day retention time and achieved 46% VS removal, whereas an identical digester fed with raw primary sludge failed at the same retention time soon after starting up. Interestingly, the %VS removal in the laboratory scale digester at 8 day retention time was greater (at 46%) than that in the full scale digester at 42 days retention time (40%), probably due to the higher operating temperature and better mixing in the laboratory scale digester.

The final sludge from the full scale digester was stable and did not undergo further fermentation. It had an earthy odour similar to that of conventional anaerobically digested sludge. The dewaterability of the final (anaerobic) sludge in terms of specific resistance to filtration was not significantly different from conventional anaerobically digested sludge.

The bacteriological analysis of the sludge indicated a 4 orders of magnitude reduction in faecal coliforms in the system and with the aerobic reactor above 50°C complete inactivation of *Ascaris* ova was achieved. The viable *Ascaris* in the digester sludge were those remaining from the seed sludge employed at the start of the investigation.

TABLE 3 BACTERIOLOGICAL QUALITY: PHASE I

Bacteriological Quality		Feed Sludge	Aerobic Sludge		Anaerobic Sludge
Faecal Coliforms†	per 100mℓ	**$9.3x10^8$**	**$7.5x10^5$**		**$4.3x10^4$**
			<50°C	>50°C	
***Ascaris* Ova**‡	(per gTS) %viable	(429) **45**	(415) **3**	(545) **0**	(758) **12**

† Median values are quoted ‡ Mean values are quoted.

Aerobic Reactor Performance

Irrespective of the presence of foam or not, the reactor operated under oxygen limiting conditions, with the result that the oxygen utilisation rate of the sludge ***(OUR)*** was limited by the oxygen transfer rate ***(OTR)*** of the aeration system.

Under non-foaming conditions, the aerobic reactor was generally operated at an air flow rate fixed at $760m^3$ (STP)/h giving an oxygen supply rate ***(OSR)*** of 1.16kg $(O_2)/m^3.h$. Under such conditions the ***OTR*** (and hence ***OUR)*** remained relatively constant at $0.15kg(O_2)/m^3.h$, representing an oxygen transfer efficiency ***(OTE)*** of 13%.

Under foaming conditions, with the air flow rate reduced to prevent spillage, a decline in the ***OTR*** and ***OUR*** was not observed due to a significant increase in the ***OTE:*** At an average ***OSR*** of $0.51kg(O_2)/m^3.h$ (less than 50% of the normal supply rate), the average ***OUR*** and ***OTE*** were $0.14kg(O_2)/m^3$ and 29% respectively, the latter more than double that obtained without foam (see Table 4).

TABLE 4 OXYGENATION CHARACTERISTICS: PHASE I

Aeration Characteristics†		No Foam	Foam
Air Flow Rate	m^3(STP)/h	**760**	**320**
OSR	$kg(O)/m^3.h$	**1.15**	**0.51**
OUR	$kg(O)/m^3.h$	**0.151**	**0.140**
OTE	%	**12.9**	**28.6**

† Mean values are quoted.

Measurement of both the oxygen and carbon dioxide concentrations in the vent gas, and assuming the nitrogen gas mass flow rate through the aerobic reactor remains constant, allowed the respiration quotient Y_{CO2} (mol CO2 produced per mol O_2 utilized) and specific heat yield Y_h (MJ heat generated per kg oxygen utilised) to be calculated. The specific heat yield and respiration quotient, calculated from the heat and oxygen mass balances, varied in the narrow range of 12.5 to 13.2 MJ/kg(O_2) and 0.68 to 0.73 mol $(CO_2)_{gen}/mol(O_2)_{ut}$ respectively under the different operating conditions, with averages of 12.9 MJ/kg(O_2) and 0.70 mol $(CO_2)_{gen}/mol(O_2)_{ut}$. These average values are virtually identical to those measured by Messenger *et al* (1993) on the Milnerton pure oxygen reactor and indicate that the measurements were accurately conducted and the materials and heat balances soundly established with the necessary sensitivity to changing operating parameters. The constancy of the Y_h and Y_{CO2} data for the different operating conditions indirectly verified the heat balance assumption that the vent gas was saturated with water vapour at all influent air flow rates. Knowing Y_h, these same equations were used to formulate a design model for the 'air' system to calculate the reactor temperature, oxygen utilisation and VS destruction rates for different operating conditions with or without foaming.

The average mass of oxygen utilised per mass of volatile solids destroyed in the aerobic reactor was 1.70kg(O_2)/kg(VS). This figure is higher than the usual value of 1.42kg(COD)/kg(VS) for sewage sludge but is in agreement with the COD/VS ratio of the influent sludge. The quantity of biological heat generated per mass of VS destroyed was calculated at 22MJ/kgVS, which shows good agreement with the recognised standard value of 21MJ/kgVS in autothermal thermophilic aerobic digestion (ATAD) obtained at laboratory scale by Andrews and Kambhu (1971). Carbon and COD mass balances of 89% and 88% respectively were achieved over the aerobic reactor.

CONCLUSIONS: PHASE I DUAL DIGESTION USING AIR

Sufficient auto-heating potential exists for the aerobic reactor, when aerated with air, to reach thermophilic temperatures throughout the year. However, because of the cooling effects of passing large volumes of nitrogen through the system, the retention times required in the reactor were relatively long (3-6 days) in comparison with the oxygen process (~1 days). The feed sludge to the reactor needs to be sufficiently concentrated so that it does not become substrate limited and therefore sludge pre-thickening should be given serious consideration. A thorough knowledge of the ***OTR-OTE*** characteristics of the aeration system is vital as this governs both the biological heat generation rate and the vent gas heat loss rates and therefore the reactor temperature. It is possible therefore that with a high transfer efficiency 'air' oxygenation device shorter retention times would be possible.

As a consequence of the long aerobic retention time, VS destruction was appreciable (25%) and the reactor tended towards an autothermal thermophilic aerobic digester. As such, the sludge may be considered to be partially stabilised rather than conditioned by aerobic treatment. Consequently, it is not considered viable to construct an air dual digester system from scratch. By extending the reactor retention time from, for example, 3-5 days to 7-8 days, the sludge could be completely stabilised (and pasteurised) in the aerobic reactor and the digester would not be required, i.e. the reactor would be designed as an ATAD.

Foaming improves the performance of the aerobic reactor by maintaining the ***OUR*** at lower air flow rates through an increase in the ***OTE***. With the reduction in vapour heat losses, higher reactor temperatures were recorded during periods of foaming. The unpredictable nature of the foaming and its apparent dependence on temperature made it difficult to fully exploit this phenomenon while maintaining control over the reactor temperature.

It was considered that by supplementing the air oxygenation system with pure oxygen, increased treatment capacity and greater temperature control of the aerobic reactor could be achieved. Further, the benefits of the pure oxygen would be retained (short retention time) at lower operating costs with a significant proportion of the oxygen demand provided by the transfer and utilisation of atmospheric oxygen from the air stream through a foam layer controlled by the air flow rate. Consequently, in 1994 the Cape Town City Council commenced with phase II of the research programme with the general aim of demonstrating the practicability of the dual digestion system employing a combination of both air and pure oxygen to improve both treatment capacity and sludge quality.

OBJECTIVES PHASE II: DUAL DIGESTION USING AIR AND OXYGEN

It was envisaged that with pure oxygen supplementation a substantial proportion of the 200m^3/d primary sludge produced at Athlone could be pasteurised and stabilised by the system without adversely affecting biogas production, which is required as fuel for a gas engine. The specific objectives of phase II were planned to cover all the claimed benefits of the dual digestion process (as with phase I) and to verify the predictions made for pure oxygen supplementation after phase I. To satisfy these objectives the following aspects of the dual digester performance were evaluated in detail:

- **The utilisation of oxygen within the aerobic reactor**; including measurement of the separate contributions from the air and pure oxygen streams.

- **The aerobic reactor heat balance**; verification of the value determined for the specific heat yield coefficient Y_h during phase I
- **The conditioning effects of aerobic pre-treatment**; As higher aerobic reactor hydraulic loading rates were possible during phase II, the retention time in the anaerobic digester could be reduced to below 10 days.
- **The performance of the anaerobic digester;** In view of the short digester retention time, the stability of the anaerobic process and its sludge product were closely monitored. Due to the high sensible heat content of the sludge transferred from the aerobic reactor, the digester operated at thermophilic temperatures.
- **The dewaterability of the final sludge**; Measured in terms of the specific resistance to filtration test.
- **The bacteriological quality of the final sludge**; Measured in terms of faecal coliform and viable *ascaris* ova concentrations.
- **The stability of the final sludge;** Measured in terms of VS and COD reduction and in terms of the residual Specific Oxygen Utilisation Rate (***SOUR***) over a fixed period of time.

OPERATION AND MONITORING: PHASE II

For phase II pure oxygen was injected into the aerobic sludge, with a Venturi type injector, at a point immediately downstream of the mixing pumps on the sludge recirculation line. This point was selected to keep the injection and subsequent utilisation of pure oxygen away from the main air stream to avoid the danger of stripping. A 55 and a 75 kW motor were fitted to the recirculation pumps to generate the necessary pressure, flow rate and discharge velocity required for the efficient transfer of pure oxygen. Operation under different flow conditions yielded valuable information with regard to defining the necessary hydraulic characteristics to achieve an acceptable oxygen transfer efficiency (***OTE***>80%). The aerobic reactor was fed with gravity thickened primary sludge (~4%TS) in batches at 1 to 2 hourly intervals, with feed sludge displacing reactor sludge to the anaerobic digester; again like in phase I not on a draw and fill basis as with most pasteurisation systems.

All parameters to allow (liquid and gas) heat and mass balances to be made across the aerobic reactor were measured with the exception of the vent gas CO_2 concentration, as a suitable instrument was no longer available. Consequently the gas mass balance was solved by accepting the measured value for the respiration quotient during phase I. The subsequent accuracy of the heat balance equations demonstrated that this approach was

correct. VS removal in both the aerobic reactor and anaerobic digester were closely monitored to determine the effect of oxygen utilisation on this parameter.

RESULTS FROM THE INVESTIGATION: PHASE II

Overall System Performance

The dual digester was monitored daily for a period of 152 days during phase II, excluding days 41 to 73 when the aerobic reactor sludge recirculation line was out of commission to allow the pipework to be modified[1]. Outside of this period the plant operated continuously without any major mechanical problems. During this time the main operating parameters such as air flow rate, pure oxygen supply rate and the influent sludge flow rate were changed to examine the influence of these parameters on the performance of the system. In all, eleven steady state periods were studied. Sufficient data was collected to fully evaluate the system. Unfortunately foaming did not occur in the reactor, and therefore could not be exploited to optimize the ***OTE*** of the air oxygenation system. A summary of plant performance during this period is provided in Tables 5-10.

TABLE 5 CHEMICAL CHARACTERISTICS PHASE II

Chemical Characteristics†		Feed Sludge	Aerobic Sludge	Anaerobic Sludge
Temperature	°C	**20** (16-26)	**57** (41-64)	**43** (33-54)
Total Solids	g/ℓ	**47** (33-67)	**39** (25-54)	**27** (9-63)
Volatile Solids	g/ℓ	**38** (26-55)	**30** (20-44)	**18** (6-34)
COD	g(O)/ℓ	**64** (42-93)	**46** (30-69)	**29** (11-61)
pH	-	**5.4** (4.8-6.2)	**7.2** (6.7-7.9)	**7.2** (6.7-7.6)
Ammonia	mg(N)/ℓ	**108** (17-200)	**518** (82-826)	**753** (480-1110)
Bicarbonate Alkalinity	mg($CaCO_3$)/ℓ	**180** (0-460)	**850** (280-1500)	**2870** (1800-3870)

† Mean values are quoted with the ranges given in brackets.

Both increased treatment capacity and greater temperature control of the aerobic reactor was achieved by increasing the oxygen transfer rate with pure oxygen injection. The average reactor retention time for the period was 1.6 days, with an average temperature of 57°C. With the aerobic reactor operating under oxygen limiting conditions, the reactor sludge temperature could be completely and instantaneously controlled by the pure

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[1] After the initial period of operation (days 1-40) it became evident that an increased pressure and flowrate in the recirculation line was necessary in order to achieve the desired pure oxygen transfer efficiency (>80%). Consequently the recirculation pipeline was modified to achieve this.

oxygen supply rate OSR^{O2}. With effective control over the reactor temperature, the shortest retention time which the reactor operated at was 0.96 days.

TABLE 6 SOLIDS REMOVAL: PHASE II

Solids Removals†		Aerobic	Anaerobic	Overall
Total Solids	%	**16**	**30**	**42**
Volatile Solids	%	**20**	**40**	**52**
COD	%	**28**	**37**	**55**

† Mean values are quoted.

On average during phase II, approximately 20% of the volatile solids in the feed sludge were removed (destroyed) in the aerobic reactor, with a further 40% removal in the anaerobic digester. The average overall percentage removal of volatile solids across the system was 52%. The higher sludge loading rates made it possible to operate the digester at short retention times (9 to 18 days). Therefore, the claim made for the dual digestion process that *aerobic pre-treatment reduces the minimum retention time for anaerobic digestion* could be tested at full scale. The high sensible heat content of the sludge from the reactor forced the digester into the thermophilic range. At the start of the evaluation period (day 1), the temperature of the sludge in the digester was as 33°C. By the end of the evaluation period (day 152) the temperature had stabilised at around 53°C at a digester retention time of 9 days. The transistion between mesophilic and thermophilic temperatures, and the decrease in retention time from 18 to 9 days, caused a temporary increase in the volatile acid alkalinity to 1000mg($CaCO_3$)/ℓ, with a concommitent decrease in pH to around 6.9 and reduction in %VS removal. However, sufficient buffer capacity was available in the feed sludge from the reactor to absorb these effects and operation in the thermophilic region was considered stable.

The conditioning effects of aerobic pre-treatment were again evident from a significant increase in the sludge ammonimum and bicarbonate alkalinity concentrations, and pH after aerobic treatment. In terms of the currently recognised measure of final sludge stability, the overall volatile solids removal efficiency of **52%** achieved is considered good and well above that considered necessary for sludge stabilisation (**>38%**, Heidman, 1989). The residual specific oxygen utilisation rate ***SOUR*** was also tested as a measure of sludge stability. Whilst the recognised standard limit of <1.0 mg(O_2)/g(TSS).h (Heidman, 1989) was not reached during the test (which was conducted over a 72 hour period), the results were significantly better than that observed for conventional mesophilic digested sludge at 20 days retention time. The dewaterability of the final (anaerobic) sludge in terms of specific resistance to filtration was not significantly

different from conventional digested sludge, although higher SRF levels were recorded; which was considered due to the exposure to higher temperatures during treatment. After undergoing secondary digestion, the dewaterability of the sludge become comparable with conventional digested sludge. During secondary digestion, it was possible to draw of the same quantities of supernatant (~**50%**) as with conventional sludge.

TABLE 7 BACTERIOLOGICAL QUALITY: PHASE II

Bacteriological Quality		Feed Sludge	Aerobic Sludge	Anaerobic Sludge
Faecal Coliforms†	per 100mℓ	$1.8x10^9$	$1.1x10^5$	$1.2x10^3$

† Median values are quoted.

The degree of disinfection provided by the system during phase II was monitored by determining faecal coliform counts at each stage in the process. An approximate 4 orders of magnitude reduction in faecal coliforms was observed across the aerobic reactor with a further 2 orders of magnitude reduction after the thermophilic anaerobic digester. With both stages in the system operating at thermophilic temperatures and the sludge flow rate at 100 m^3/d the concentration of faecal coliforms in the both the aerobic and final sludge decreased to $<10^3$ /100mℓ. However, short circuiting was evident when the sludge flow rate was increased to 196m^3/d, with counts in the range 10^4 to 10^6/100mℓ for the aerobic sludge and ~10^4/100mℓ for the final sludge being observed. It was concluded, that to provide effective pasteurisation, feeding on a draw and fill basis would be required.

Aerobic Reactor Performance

Throughout phase II the aerobic reactor operated under oxygen limiting conditions, with the result that the oxygen utilisation rate of the sludge ***OUR*** was limited by the separate contributions to the overall oxygen transfer rate OTR^{AIR+O2} from the pure oxygen and air oxygenation systems (OTR^{AIR} + OTR^{O2}). The independence the oxygenation systems was verified by the fact that (1) the 'pure oxygen' tansfer efficiency OTE^{O2} remained constant whether or not the air supply was on or off, and (2) the oxygen transfer efficiency of oxygen from the air stream OTE^{AIR} did not vary significantly from an average value of 11.9% in spite of changes made to the pure oxygen supply rate. With the air and pure oxygen streams acting independently of each other, the oxygen utilisation rate ***OUR*** could likewise be considered to consist of separate contributions from the oxygen derived from the *air* stream and from *pure oxygen* injection (i.e. ***OUR*** = OUR^{AIR} + OUR^{O2}).

TABLE 8 OXYGENATION CHARACTERISTICS: PHASE II

No.	$Q(AIR)_{in}$ $m^3(STP)/h$	$M(O_2)_{in}^{O2}$ kgO/h	OSR^{AIR} $kgO/m^3.h$	OSR^{O2} $kgO/m^3.h$	OSR^{AIR+O2} $kgO/m^3.h$	OUR^{AIR} $kgO/m^3.h$	OUR^{O2} $kgO/m^3.h$	OUR^{AIR+O2} $kgO/m^3.h$	OTE^{AIR} %	OTE^{O2} %
1	260	60	0.395	0.326	0.721	0.047	0.151	0.198	11.8	46.3
2	610	60	0.919	0.326	1.245	0.113	0.156	0.269	12.3	47.9
3	610	60	0.925	0.326	1.252	0.108	0.178	0.286	11.7	54.6
4	0	60	0	0.326	0.326	0	0.180	0.180	0	55.2
5	0	24	0	0.130	0.130	0	0.108	0.108	0	83.0
6	380	24	0.572	0.130	0.702	0.068	0.108	0.176	11.8	83.0
7	680	24	1.025	0.130	1.155	0.121	0.109	0.230	11.8	83.5
8	0	36	0	0.196	0.196	0	0.162	0.162	0	82.6
9	320	60	0.483	0.326	0.809	0.058	0.266	0.324	12.1	81.5
10	360	96	0.537	0.522	1.059	0.063	0.435	0.498	11.7	83.3
11	420	76	0.628	0.413	1.041	0.076	0.341	0.417	12.1	82.6

The maximum oxygen transfer rate OTR_{max}^{O2} effected by the pure oxygen system during the evaluation period was 0.435 $kg(O_2)/m^3.h$, which at an transfer efficiency OTE^{O2} of 83.3% corresponds to an oxygen supply rate OSR^{O2} of 0.537 $kg(O_2)/m^3.h$. The system was not tested to determine if significantly higher OTR^{O2} rates than 0.435 $kg(O_2)/m^3.h$ could have been achieved as this would have resulted in the system overheating. The limitations on oxygen solubility under the hydraulic conditions in the recirculation pipeline would suggest however that the effective maximum OTR_{max}^{O2} is not much higher than 0.435 $kg(O_2)/m^3.h$ recorded.

The maximum biological oxygen utilisation rate ***OUR*** recorded during the evaluation period was 0.498 $m^3/kg(O_2).h$ (where OUR^{AIR}=0.063 and OUR^{O2}=0.435). At this time the average total solids concentration of the primary sludge was 42kg(TS)/m^3 and the retention time in the aerobic reactor was 0.96 days. The fact that the reactor remained oxygen limiting at this high ***OUR*** level demonstrates that the aerobic process could support higher ***OUR***'s up to the point where OUR^{max} is reached. An estimate of this OUR^{max} (when the reactor becomes substrate limited) for different sludge types is made from an application of a kinetic model which was developed to describe the rate of enzyme hydrolysis, and consequently the rate of availability of readily biodegradable volatile solids that provide substrate for the biological thermophilic reactions.

During phase II, the specific heat yield Y_h was measured to be 12.8 $MJ/kg(O_2)$. This is the same value as that determined during phase I and shows good agreement with the value of 13.0 obtained by Messenger *et al* (1993). The spread in Y_h values between each steady state period was small (12.3 to 13.5 $MJ/kg(O_2)$) in spite of the widely differing operating conditions during the 11 steady state periods. The linear relationship between the biological heating rate H_b and the ***OUR*** is shown graphically in Figure 2, for both

phase I and phase II data. The close correlation between H_b and ***OUR*** confirms the accuracy of the mass and heat balance equations and tests carried out at Athlone under the wide range of operating conditions using air, pure oxygen and a combination of these. It also confirmed the accuracy of the value of Y_{co2} =0.70 mol$(CO_2)_{gen}$/mol$(O_2)_{ut}$ for the respiration quotient accepted from phase I.

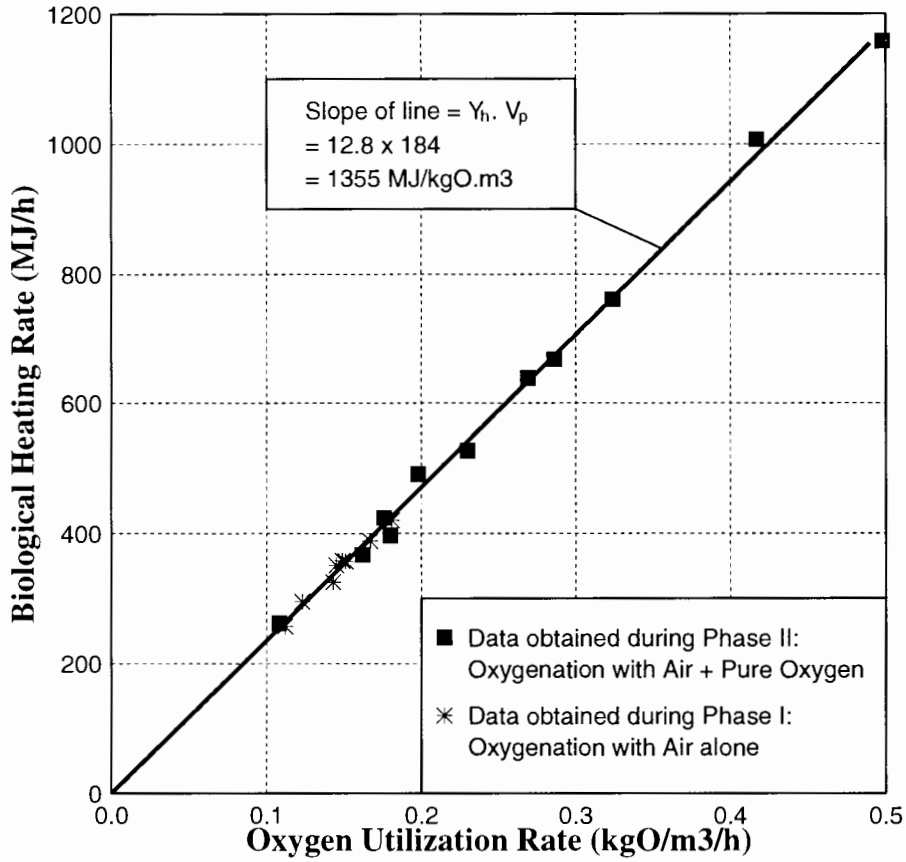

FIGURE 2 BIOLOGICAL HEATING RATE VERSUS *OUR*

As in phase I, the rate of volatile solids destruction in the aerobic reactor could be linked to the oxygen utilisation rate ***OUR***. This relationship, closely equal to the COD/VS ratio of the raw feed sludge (**1.7** kg(O)/kg(VS)), was maintained over the full range of operating conditions at Athlone (retention time 1 to 4 days, oxygenation with air and/or pure oxygen). The quantity of biological heat generated per mass of VS destroyed was calculated at **19** MJ/kg(VS) which shows reasonable agreement with the 22MJ/kg(VS) obtained in phase I and the 21MJ/kg(VS) obtained by Andrews and Kambhu (1971) for ATAD systems applications.

MODELLING THE DUAL DIGESTION PROCESS

Sufficient information was obtained in phases I and II of this investigation to allow a mathematical model for dual digestion to be compiled which can reliably simulate all the main operating parameters of the system with air and/or pure oxygen oxygenation

and meso- or thermophilic anaerobic digestion. In addition to those components in the model calibrated with operational data from the Athlone plant, a number of others were incorporated based on data and kinetics from other sources, namely:

(1) **The onset of substrate limitation in the aerobic reactor** is predicted using a kinetic model (based on the kinetic model of Dold *et al*, 1991) which describes the rate of enzyme hydrolysis, and therefore the rate of availability of readily biodegradable volatile solids to feed the oxidative reactions;

(2) **The rate of volatile solids destruction and biogas production in the digester** is predicted using a kinetic equation developed by Grau *et al* (1975) and used to describe the rate of conversion of acetate to methane, which is considered to be the rate limiting step. The stoichiometry describing the breakdown of VS through the different metabolic pathways prior to acetate conversion (described by Guger *et al* , 1993) follows that of Siegrist *et al* (1993);

(3) **The anaerobic digester operating temperature** is predicted from a steady state heat balance for the digester calibrated with actual operating data from four full scale plants.

In using the model, two approaches can be taken, either to (1) evaluate system performance for a given aerobic reactor and anaerobic digester retention time or (2) calculate retention time of the reactor and/or digester for a defined system performance i.e. sludge quality. Different dual digester system configurations can be analysed and comparison can be made with conventional anaerobic digestion (mesophilic or thermophilic). The model includes an approximate cost analysis for both operating and capital costs. The information provided by the model is as follows:

- Oxygenation characteristics of the aerobic reactor (for air and/or pure oxygen)
- Steady state heat balance for the aerobic reactor
- Heating requirements for the anaerobic digester
- Effect of installing heat exchangers
- Volatile solids destruction in both the reactor and digester.
- Biogas production in the digester
- Performance of an installed gas engine
- The stability of the final sludge.
- Minimum reactor retention time to prevent substrate limitation.
- Minimum digester retention time to ensure sludge stability.

- Approximate system operating and capital costs with or without interstage or afterstage heat exchange or external heat via a boiler or gas engine..

A number of simulations with the model were made which yielded information with regard to:

- The effect of ambient temperature on system VS removal, biogas production and the required oxygen utilisation rate.

- The effect of increasing the proportion of external heat to the aerobic reactor on system VS removal, biogas production and the oxygen supply rate.

- The effect of improved oxygen transfer efficiency of the air oxygenation device on the rate of pure oxygen injection.

- Prediction of the minimum retention time to prevent substrate limitation, for four different feed sludge types (viz primary, waste activated and mixtures of primary and humus, and primary and waste activated sludges) for different feed sludge solids concentrations and ambient (feed sludge and surrounding air) temperatures.

CONCLUSIONS

The dual digestion system proved relatively simple to operate and is capable of producing a **pasteurised** and well **stabilised** sludge product at short anaerobic digester retention times. The claimed benefits of the system were all verified with the exception of the claim that the aerobic reactor heat pre-treatment of the sludge allows the anaerobic digester to operate at much shorter retention times (~10 days). However, the digester can be operated at 10 days retention time provided it is in the thermophilic temperature range, in which case an adequately stabilised sludge product will be produced. If the digester is operated in the mesophilic range, even though the digester process per se is stable at 10 days retention time, the sludge product is not stable and has a high residual biodegradable VS content. Consequently, it is not so much the stability of the anaerobic process that governs the digester retention time as has been suggested in the past, but rather the stability of the final sludge product. In the mesophilic range a digester retention time of at least 15 days or longer is required to produce a sludge product of equivalent stability as conventional mesophilic anaerobic digestion at 20 days retention time.

When treating primary sludge (provided the sludge is sufficiently concentrated, >3%TS) with pure oxygen, the reactor sludge temperature can be completely and instantaneously controlled, and aerobic retention times as short as 1 day are practically possible. If the sludge contains a significant fraction of secondary sludge, longer retention times are required due to the shortage of available biodegradable substrate. The principal operating parameters for the system can be predicted using the dual digestion simulation model which was developed.

Where the installation, or expansion, of a conventional anaerobic digestion plant is considered, it is recommended that a technical and economic feasibility evaluation be done with the simulation to determine the cost-effectiveness and performance of the dual digestion system. ***Dual digestion may produce a better quality product at a cheaper overall cost.***

The upgrading of a conventional anaerobic digestion plant to dual digestion is considered a relatively simple task involving the addition of an appropriately sized aerobic reactor with the necessary set of mechanical equipment to drive the process. The characteristics of the oxygenation device, which need to be well defined, must be capable of effecting a high oxygen transfer efficiency. For pure oxygen this ensures that gas wastage is minimised. For air, through a reduction in the vent gas heat losses, reactor retention times as short as 1 day may be possible, although the type of air oxygenation device required would need to be similar to the method of pure oxygen injection i.e. to inject the air into the recirculation line, under pressure; a method employed by Wolinski (1985) who reported 'air' oxygen transfer efficiencies close to 100% during foaming.

Consideration must be given to the interdependence of the operating temperatures of the reactor and digester, otherwise there is a high risk of the anaerobic digester temperature falling between the mesophilic and thermophilic temperature regions. With the high ambient temperatures encountered in South Africa, for short digester retention times (10 to 15 days), interstage cooling is essential for the digester to be operated in the **mesophilic** range. Alternatively, the digester can be allowed to operate in the **thermophilic** range by controlling the aerobic reactor temperature to accommodate seasonal temperature changes.

Ahring (1994) recommends an operating temperature in the range 51° to 55°C for thermophilic digestion and considers that with adequate temperature control, the thermophilic process is as stable as its mesophilic counterpart.The benefit of thermophilic anaerobic treatment is that the sludge is fully stabilised and pasteurised at 10 days

retention time. Perhaps the principal reason why the stand alone thermophilic digestion process has not become preferable to mesophilic digestion is the susceptibility of the thermophilic digestion process to upset with temperature fluctuations. Dependence on a single heating source subject to sporadic failure is a likely source for thermophilic process failure. The thermophilic aerobic reactor is an appropriate pre-treatment system for thermophilic digestion providing a buffer for both temperature and pH changes in the digester. If the option of operating the digester in the thermophilic region is selected to benefit from its short retention time, then it is recommended that the biological heating of the aerobic reactor be supplemented with a secondary external heat source. In addition to reducing the required rate of pure oxygen injection necessary for specific reactor temperatures, both heating systems provide back up for each other in the event of failure, allowing thermophilic temperatures to be sustained. The two sources of external heat are:

(1) the recovery of sensible heat from either the hot aerobic sludge (interstage heat exchange) or the hot digester sludge (afterstage heat exchange)

(2) the generation of heat from the burning of the biogas generated during anaerobic digestion. Typically either through the use of conventional boilers (hot water systems) or the recovery of heat from an installed gas engine.

From a performance evaluation with the dual digestion simulation model of various dual digester configurations with or without the various external heat source options and comparing these with conventional mesophilic and thermophilic digestion it was concluded that the most technically and economically feasible configuration is that shown in Figure 3 i.e. dual digestion with thermophilic digestion generating gas for a gas engine from which heat is recovered to supplement the biologically generated heat in the aerobic reactor. While the biological heat can be generated with liquid oxygen, additional cost savings (up to 50%) on oxygen supply can be made by on-site oxygen generation by a Vacuum Swing Adsorption (VSA) plant. The VSA plant is ideal for application to this dual digestion configuration which requires a constant 24 hour oxygen output. The additional heat requirements which arise from seasonal ambient (feed sludge and surrounding air) temperature fluctuations can be met by the gas engine and heat exchanger external heat source. This dual digestion system configuration compares very favourably both technically and economically with conventional mesophilic digestion and produces a superior sludge product which can be beneficially used in agriculture.

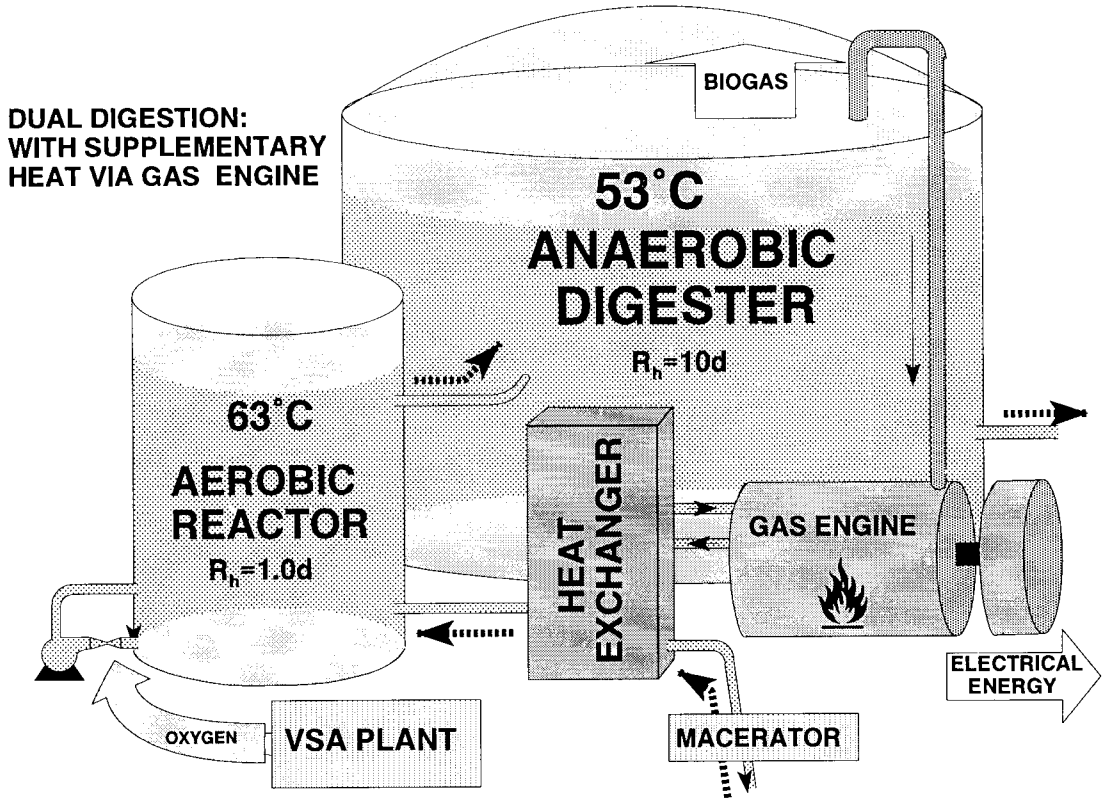

FIGURE 3 DUAL DIGESTION SYSTEM CONFIGURATION INCORPORATING HEAT RECOVERY FROM AN INSTALLED GAS ENGINE

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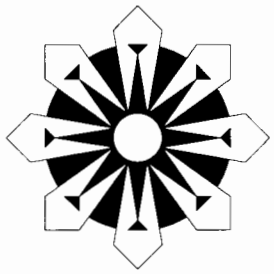

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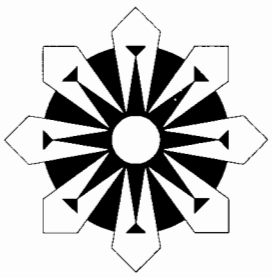

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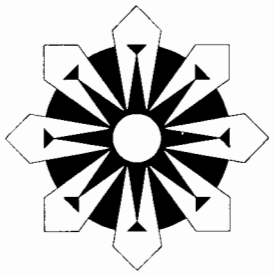

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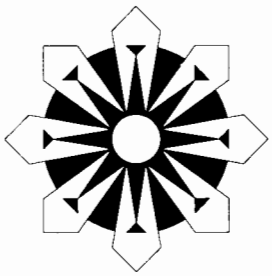


LIST OF SYMBOLS

A	=	cross-sectional area at right angles to the flow of heat through a wall (m^2)
$\%(Ar)_{in}$	=	percent volumetric concentration of argon in the influent gas stream (% v/v)
ATAD	=	autothermal thermophilic aerobic digestion process
$BA_{in,r,d}$	=	bicarbonate alkalinity concentration in the influent (raw), aerobic reactor, and anaerobic digester sludges ($mg(CaCO_3)/l$)
$[BVS]_{in,out}$	=	biodegradable volatile solids concentration in the influent and effluent sludge from the aerobic reactor ($kg(BVS)/m^3$)
BVS	=	biodegradable volatile solids
b	=	organism decay rate (/d)
$\%(CO_2)_{in,out}$	=	percent volumetric concentration of carbon dioxide in the influent and effluent gas stream (% v/v)
$C(O_2)_{P,T}^{sat}$	=	oxygen solubility at pressure P and temperature T ($kg(O_2)/m^3$)
$COD_{in,r,d}$	=	chemical oxygen demand in the influent (raw), aerobic reactor, and anaerobic digester sludges (g(O)/l)
$COND_{in,r,d}$	=	conductivity in the influent (raw), aerobic reactor, and anaerobic digester sludges (mS/m)
COSΦ	=	power factor for the mixing device, defined as the ratio of the power consumption to the volt-amperes, Φ is the angular phase difference between the voltage and current
$C_p^{20°C}(AIR)$	=	heat capacity of dry atmospheric air at 20°C (MJ/kg.°C)
$C_p\ (AIR)_{in,out}$	=	heat capacity of the dry influent and effluent air streams respectively from the aerobic reactor (MJ/kg.°C)
$C_p^{20°C}(SL)$	=	heat capacity of sewage sludge (estimated) at 20°C ($MJ/m^3.°C$)
$C_p^{50°C}(H_2O)_{vap}$	=	heat capacity of water vapour at 50°C (MJ/kg.°C)
$\%CH_4$	=	concentration of methane in the biogas from the digester (%)
$\%CO_2$	=	concentration of carbon dioxide in the biogas from the digester (%)
D	=	diameter of the recirculation line pump impeller (mm)
ΔNH_4	=	increase in ammonium ion concentration after aerobic treatment ($g(N)/m^3$)
DNH&PD	=	Department of National Health and Population Development
e	=	emmissivity of a grey body
f_{bio}	=	fraction of the volatile solids in the influent sludge to the aerobic reactor which are biodegradable (kg(BVS)/kg(VS))
f_e	=	endogenous residue fraction (kg(COD)/kg(COD))
$FC_{in,r,d}$	=	faecal coliform concentration in the influent (raw), aerobic reactor, and anaerobic digester sludges (/cl)

f_{mech} = fraction of electrical power drawn by the electrical motor on the mixing device which is converted to heat in the reactor sludge (-)

f_n = fraction of nitrogen in the influent sludge volatile solids (g(N)/kg(VS))

f_{ovs} = mass of oxygen utilised per mass of volatile solids destroyed in the aerobic reactor (kg(O_2)/kg(VS))

f_{vs} = fraction of the total solids in the influent sludge to the aerobic reactor which are volatile (kg(VS)/kg(TS))

f_{bvs} = fraction of the volatile solids in the influent sludge to the aerobic reactor which are biodegradable (kg(BVS)/kg(VS))

f_{rbvs} = fraction of the biodegradable volatile solids in the influent sludge to the aerobic reactor which are readily biodegradable (kg(RBVS)/kg(BVS))

$f(VS)_{rem}^{O2lim}$ = fraction of volatile solids removed in the aerobic reactor under oxygen limiting conditions (-)

$f(VS)_{rem}^{VSlim}$ = fraction of volatile solids removed in the aerobic reactor under substrate limiting conditions (-)

$\boldsymbol{GR}_j$ = generation rate of component j in the anaerobic process (kg(COD)/m^3.h)

$\boldsymbol{H}_g$ = total net rate of sensible heat loss from the aerobic reactor with the effluent gas stream (MJ/h)

$\boldsymbol{H}_g(AIR)$ = net rate of sensible heat loss from the aerobic reactor attributable to the dry air component in the effluent gas stream (MJ/h)

$\boldsymbol{H}_g(H_2O)$ = net rate of sensible heat loss from the aerobic react or attributable to the water vapour in the effluent gas stream (MJ/h)

$\boldsymbol{H}_b$ = rate of biological heat generation within the aerobic reactor (MJ/h)

$\boldsymbol{H}_b^{AIR}$ = rate of biological heat generation attributed to oxygen utilisation from the air stream when considering pure oxygen supplementation (MJ/h)

$\boldsymbol{H}_{b\ max}$ = maximum rate of biological heat generation within the aerobic reactor, determined by the maximum biological oxygen utilisation rate under substrate sufficient conditions (MJ/h)

$\boldsymbol{H}_b^{O2}$ = rate of biological heat generation attributed to oxygen utilisation from the pure oxygen injection stream when considering pure oxygen supplementation (MJ/h)

$\boldsymbol{H}_m$ = rate of mechanical heat energy input to the aerobic reactor from the mixing device (MJ/h)

$\boldsymbol{H}_s^*$ = total net rate of sensible heat loss from the aerobic reactor with the effluent sludge stream ignoring the effects of water vaporisation (MJ/h)

$\boldsymbol{H}_s$ = total net rate of sensible heat loss from the aerobic reactor with the effluent sludge stream (MJ/h)

$\boldsymbol{H}_v$ = total net rate of vapour heat loss from the aerobic reactor with the effluent gas stream (MJ/h)

$\boldsymbol{H}_w$ = total rate of heat loss through the walls of the aerobic reactor (MJ/h)

$\boldsymbol{h}$ = heat transfer coefficient for convection (MJ/m^2.h.°C)

$\boldsymbol{\delta H}_p$ = change in heat content of a substance at constant pressure (MJ)

h_f = head loss due to friction in the reactor sludge recirculation line (m)
transfer coefficient for radiation (MJ/m^2.h.°C)

h_r = heat transfer coefficient for radiation (MJ/m^2.h.°C)

$H_w^{a,b,c,d,e,f}$ = rate of heat loss through portions a,b,c,d,e, and f of the walls of the aerobic reactor respectively, see Figure 3.8 (MJ/h)

I = phase current drawn by the electrical motor driving the mixing device for the aerobic reactor (Amps)

K_{MP} = heterotroph maximum specific growth rate on SBCOD (/h)

K_{SP} = half saturation constant for SBCOD (kg(COD)/kg(COD)

k_d = volatile solids destruction rate coefficient; employed in describing the kinetics of VS removal under substrate limiting conditions (/d)

k = thermal conductivity of a specific substance (MJ/h.m.°C)

L_v = latent heat of vaporisation of water (MJ/kg)

$L_v^{20°C}$ = latent heat of vaporisation of water at 20°C (MJ/kg)

$L_v^{50°C}$ = latent heat of vaporisation of water at 50°C (MJ/kg)

$M(AIR)_{in,out}$ = influent and effluent dry air mass flow rates (kg/h)

$M(CO_2)_{gen}$ = mass rate of carbon dioxide generation within the aerobic reactor (kg(CO_2)/h)

$M(CO_2)_{in,out}$ = mass rate of carbon dioxide entering and leaving the aerobic reactor in the gas stream (kg(CO_2)/h)

$M(GAS)_{in,out}$ = influent and effluent air mass flow rates, including the water vapour (kg/h)

$M(H_2O)_{in,out}$ = influent and effluent mass water vapour flow rates (kg(H_2O)/h)

$M(H_2O)_{vap}$ = mass rate of water vaporisation from the sludge in the reactor (kg(H_2O)/h)

$M(NH_4^+ - N)_{gen}$ = mass rate of ammoniacal nitrogen generated in the aerobic reactor through the breakdown of volatile solids (g(NH_4^+ - N)/h)

$M(NH_4^+ - N)_{in,out}$ = mass rate of ammoniacal nitrogen entering and leaving the aerobic reactor in the influent & effluent sludges respectively(g(NH_4^+ N)/h)

$M(N_2)_{in,out}$ = mass rate of nitrogen entering and leaving the aerobic reactor in the gas stream (kg(N_2)/h)

$M(O_2)_{in,out}$ = mass rate of oxygen entering and leaving the aerobic reactor in the gas stream (kg(O_2)/h)

$M(O_2)_{sup}$ = mass rate of pure oxygen supplementation into the aerobic reactor (kg(O_2)/h)

$M(O_2)_{trans}^{max}$ = maximum oxygen transfer rate which can be effected by the oxygenation device (kg(O_2)/h)

$M(O_2)_{ut}$ = mass rate of oxygen utilisation within the aerobic reactor (kg(O_2)/h)

$M(VS)_{dest}$ = mass rate of volatile solids destruction in the aerobic reactor (kg(VS)/h)

$M(VS)_{in,r}$ = mass rate of volatile solids entering and leaving the aerobic reactor in the influent and effluent sludges respectively (kg(VS)/h)

$m_w(AIR)_{in,out}$ = molecular mass of the dry air in the influent and effluent gas streams (kg/kmol)

$m_w(CO_2)$ = molecular mass of gaseous carbon dioxide (kg/kmol)

$m_w(H_2O)$ = molecular mass of water (kg/kmol)

$m_w(N_2)$ = molecular mass of equivalent nitrogen i.e. atmospheric nitrogen which incorporates argon (kg/kmol)

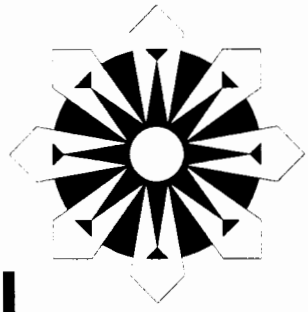
$m_w(O_2)$ = molecular mass of gaseous oxygen (kg/kmol)

$\hat{\mu}$	=	organism maximum specific growth rate (/d)
μ_{JT}	=	Joule-Thompson Coefficient, describing the change in temperature of a gas as it passes from a region of higher pressure into a region of lower pressure where the expansion of the gas is adiabatic. (°C/atm)
$\%(N_2)_{in,out}$	=	percent volumetric concentration of equivalent nitrogen (inc. argon) in the influent and effluent gas stream (% v/v)
N	=	rotational speed of the pump motor in the sludge recirculation line (rpm)
$n(AIR)_{in,out}$	=	influent and effluent dry air molar mass flow rates (kmol/h)
$n(CO_2)_{gen}$	=	molar mass rate of carbon dioxide generation within the aerobic reactor (kmol/h)
$n(CO_2)_{in,out}$	=	molar mass flow rate of carbon dioxide entering and leaving the aerobic reactor in the gas stream (kmol/h)
$n(GAS)_{in,out}$	=	influent and effluent air molar mass flow rates, including the water vapour (kmol/h)
$n(H_2O)_{in,out}$	=	influent and effluent molar mass water vapour flow rates (kmol/h)
$n(H_2O)_{vap}$	=	molar mass rate of water vaporisation from the sludge in the reactor (kmol/h)
$NH^+_{4in,r,d}$	=	ammoniacal nitrogen concentration in the influent (raw), aerobic reactor, and anaerobic digester sludges (mg(N)/l)
$n(N_2)_{in,out}$	=	molar mass flow rate of nitrogen entering and leaving the aerobic reactor in the gas stream (kmol/h)
$n(O_2)_{ut}$	=	molar mass rate of oxygen utilisation within the aerobic reactor (kmol/h)
$n(O_2)_{in,out}$	=	molar mass flow rate of oxygen entering and leaving the aerobic reactor in the gas stream (kmol/h)
$\%(O_2)_{in,out}$	=	percent volumetric concentration of oxygen in the influent and effluent gas stream (% v/v)
OLR	=	organic loading rate to the anaerobic digester (kg(COD)/m^3)
OR_j	=	output rate of component j in the anaerobic process (kg(COD)/m^3.h)
OSR	=	volume specific mass oxygen supply rate to the aerobic reactor (kg(O_2)/m^3.h)
OTE	=	oxygen transfer efficiency i.e. the fraction of oxygen supplied to the reactor in the air stream which the bacteria in the sludge are able to utilise (-)
OTR	=	volume specific mass oxygen transfer rate i.e. the rate at which oxygen is transferred to the sludge by the aeration device (kg(O_2)/m^3.h)
OTR_{max}	=	maximum volume specific mass oxygen transfer rate i.e. the maximum rate at which oxygen can be transferred to the sludge by the aeration device (kg(O_2)/m^3.h)
OTR^{std}_{max}	=	standard maximum volume specific mass oxygen transfer rate i.e. the maximum rate at which oxygen can be transferred to the sludge by the aeration device under standard conditions (kg(O_2)/m^3.h)
OUR	=	volume specific mass biological oxygen utilisation rate i.e. the rate at which the bacteria in the sludge utilise the oxygen transferred into solution (kg(O_2)/m^3.h)

OUR^{AIR}	=	volume specific mass biological oxygen utilisation rate which can be attributed to the oxygen transferred from the air stream when pure oxygen supplementation is employed ($kg(O_2)/m^3.h$)
OUR_{max}	=	volume specific maximum mass biological oxygen utilisation rate i.e. the maximum rate at which the bacteria, under oxygen sufficient conditions, can utilise the oxygen transferred into solution ($kg(O_2)/m^3.h$)
OUR^{O2}	=	volume specific mass biological oxygen utilisation rate which can be attributed to the oxygen transferred from the pure oxygen injection stream when pure oxygen supplementation is employed ($kg(O_2)/m^3.h$)
OTE^{O2}	=	pure oxygen transfer efficiency i.e. the fraction of pure oxygen supplied to the reactor during pure oxygen supplementation which the bacteria in the sludge are able to utilise (-)
$P(AIR)_{in,out}$	=	partial pressure exerted by the dry air component in the aerobic reactor influent and effluent gas streams respectively (mmHg)
dP^{sat}/dT	=	rate of change of saturation pressure with temperature for a two phase (liquid-gas) system in equilibrium in terms of mass and heat transfer ($MJ/m^3.K$)
P	=	electrical power drawn by the motor driving the mixing device for the aerobic reactor (kW)
$P(GAS)_{in,out}$	=	total pressure exerted by the aerobic reactor influent and effluent gas streams respectively, dry air + water vapour(mmHg)
$P(H_2O)_{in,out}$	=	partial pressure exerted by the water vapour in the aerobic reactor influent and effluent gas streams respectively (mmHg)
$P(H_2O)^{sat}_{in,out}$	=	saturation partial pressure exerted by the water vapour in the aerobic reactor influent and effluent gas streams respectively (mmHg)
$P(O_2)$	=	oxygen partial pressure at the point of injection into the sludge recirculation line (mmHg)
$pH_{in,r,d}$	=	pH in the influent (raw), aerobic reactor, and anaerobic digester sludges (-)
P^{sat}	=	saturation partial pressure of a substance at temperature **T** (mmHg)
P^{sat}_0	=	saturation partial pressure of a substance at reference temperature $\mathbf{T}_0$ (mmHg)
$Q(AIR)_{in,out}$	=	influent and effluent dry air volumetric flow rates (m^3(STP)/h)
δq	=	quantity of heat transferred to a specific substance which brings about a temperature rise of δt(MJ)
Q	=	rate of heat transfer either onto, or through, or away from a specific wall area (MJ/h)
$Q(GAS)_{in,out}$	=	influent and effluent air volumetric flow rates, including the water vapour (m^3(STP)/h)
$Q(H_2O)_{vap}$	=	rate of volume displacement by vaporisation from the sludge in the aerobic reactor(m^3/h)
$Q(SL)_d$	=	effluent volumetric sludge flow rate from the anaerobic digester (m^3/d)
$Q(SL)_{in,out}$	=	influent and effluent volumetric sludge flow rates to and from the aerobic reactor(m^3/d)
$Q(SL)_{mix}$	=	volumetric sludge flow rate in the sludge recirculation line (m^3/h)

RBVS	=	readily biodegradable volatile solids
R_e	=	Reynolds number (-)
R_0	=	ideal gas constant (mmHg.m^3/kmol.K)
R_h	=	hydraulic retention time in the aerobic reactor (d)
R_h^{max}	=	maximum sludge retention time at which the aerobic reactor can operatebefore it becomes substrate limited (d)
σ	=	Stefan-Boltzmann constant (MJ/m^2.h.K^4)
S	=	effluent substrate concentration (kg(COD)/m^3)
S_0	=	influent substrate concentration (kg(COD)/m^3)
SBCOD	=	slowly biodegradable particulate COD
$SOUR$	=	specific oxygen utilisation rate which the final sludge stimulates under laboratory batch test conditions (g(O_2)/kg(TSS).h)
SRF_d	=	specific resistance to filtration of the effluent sludge from the anaerobic digester (m/kg x 10^{12})
$T(AIR)_{in,out}$	=	temperature of the influent and effluent air stream from the aerobic reactor (°C)
$T(AIR)_r$	=	temperature of the air stream as it leaves the surface of the sludge in the aerobic reactor i.e. in the aerobic reactor head space (°C)
$T(AMB)$	=	ambient temperature (°C)
δT	=	temperature rise effected by the transfer of heat #deltaq# to a specific substance (°C)
$T(FM)_r$	=	temperature of the foam in the aerobic reactor (°C)
$T(GAS)_{dig}$	=	temperature of the biogas in the anaerobic digester head space (°C)
$T(GAS)_{in,out}$	=	absolute temperture of the influent and effluent aerobic reactor gas streams respectively, dry air + water vapour (°C)
$T(GAS)_{in,out}$	=	temperature of a gas on the inside and outside respectively of a wall, employed when calculating the rate of heat loss through the wall (°C)
TS	=	total solids
TS_{in}^{min}	=	minimum influent sludge total solids concentration required to prevent substrate limitation from taking place in the aerobic reactor (kg(TS)/m^3)
$TS_{in,r,d}$	=	total solids concentration in the influent (raw), aerobic reactor, and anaerobic digester sludges (g/l)
$T(SL)_{in,r,d}$	=	temperature of the influent (raw), aerobic reactor, and anaerobic digester sludges (°C)
$T(W)_{in,out}$	=	inside and outside temperature respectively of a wall employed when calculating the rate of heat loss through the wall (°C)
$U_{in}^{20°C}$	=	humidity of the influent aerobic reactor gas stream at 20 C assuming that the gas is fully saturated with water vapour (kg(H_2O)/kg(AIR))
$U_{in}^{50°C}$	=	humidity of the effluent aerobic reactor gas stream at 50 C assuming that the gas is fully saturated with water vapour (kg(H_2O)/kg(AIR))
$U_{in,out}$	=	humidity of the influent and effluent aerobic reactor gas streams respectively (kg(H_2O)/kg(AIR))
$VA_{in,r,d}$	=	volatile acid alkalinity concentration in the influent (raw), aerobic reactor, and anaerobic digester sludges (mg$CaCO_3$/l)

V	=	phase voltage across the electrical motor driving the mixing device for the aerobic reactor (kV)
$V_G - V_L$	=	change in volume accompanying the phase change for a two phase (liquid-gas) system in equilibrium in terms of mass and heat transfer (m^3)
$VI_{in,r,d}$	=	fraction of viable ascaris ova as a percentage of the total ascaris ova count (%)
V_p	=	liquid process volume of the aerobic reactor, excluding the foam space (m^3)
$VS_{in,r,d}$	=	volatile solids concentration in the influent (raw), aerobic reactor, and anaerobic digester sludges (g/l)
VS	=	volatile solids
x	=	thickness of a wall, employed when calculating the rate of heat loss through the wall (m)
X	=	anaerobic active organism concentration ($kg(COD)/m^3$)
Y'_{CO2}	=	respiration quotient i.e. the mass of carbon dioxide generated per unit mass of oxygen utilised by the bacteria in the sludge in the aerobic reactor ($kg(CO_2)/kg(O_2)$)
Y_{CO2}	=	respiration coefficient i.e. the molar mass of carbon dioxide generated per unit molar mass of oxygen utilised by the bacteria in the sludge in the aerobic reactor ($mol(CO_2)_{gen}/mol(O_2)_{ut}$)
Y_h	=	specific heat yield coefficient; the quantity of biological heat generated per unit mass of oxygen utilised ($MJ/kg(O_2)$)
Y_{ZH}	=	heterotroph biomass yield ($kg(COD)/kg(COD)$)
Z_{BH}	=	heterotroph active biomass concentration ($kg(COD)/m^3$)



CHAPTER 1

INTRODUCTION

1.1 MUNICIPAL WASTEWATER TREATMENT

1.1.1 Conventional Treatment Practice

The objective of municipal wastewater treatment may be identified generally as the treatment of the collected wastewater to an effluent quality standard acceptable for discharge to the environment. During conventional treatment settleable raw wastewater solids are removed from the water phase by primary sedimentation and then concentrated to form primary sludge. The remaining non-settleable solids are degraded and converted in biological reactor systems into settleable biomass solids, which in the secondary sedimentation stage is removed from the treated wastewater to produce a clear water stream as effluent. In the biological treatment system, whether trickling filter or activated sludge, biomass solids are continuously generated and surplus biomass needs to be harvested from the system to control the biomass accumulation in the system. The biomass solids harvested from the system are concentrated to form secondary sludge.

Having treated the wastewater to an acceptable quality, it remains to treat and dispose of the primary and secondary sludges produced in a way that safeguards the environment against pollution and the community against disease. Stabilisation is also required from the point of view of reducing the oxygen demand (or energy content) of the sludge to produce an essentially unbiodegradable organic (and inorganic) humus. The treatment and disposal of the primary and secondary sludge, which together is called sewage sludge, is one of the major challenges facing urban municipalities today and is usually more burdensome economically and troublesome technically than treating the wastewater that produced it.

1.1.2 Sludge Treatment

The sewage sludges produced during wastewater treatment are essentially concentrated aqueous suspensions of putrefactive particulate organic matter of which a large fraction is biodegradable. Present in the sludges are a large number of different pathogens, the level and variety of which are dependent upon the state of health of the community. In addition, if there is a significant industrial contribution to the municipal wastewater, the sludge may contain high levels of heavy metals and other toxic chemicals.

The necessity for treating sludge prior to disposal has therefore long been recognised. The putrefactive nature of sludge makes it a potential public nuisance and health hazard as uncontrolled biodegradation of sludge rapidly generates unpleasant odours and provides sites for insect and fly breeding. Indiscriminate disposal of the sludge can result in the pollution of surface and ground waters and contaminate land with pathogens and heavy metals.

Historically, treatment has been aimed at odour control which requires reducing the organic content of the sludge by controlled degradation of the putrefactive organic material. This process is termed **stabilisation**. As disposal invariably involves transporting the stabilised sludge to a place where its presence is not objectionable, sludge treatment is aimed also at reducing the water content of the stabilised sludge. This process is termed **dewatering** and makes transportation and disposal more economical.

1.1.3 Sludge Disposal Options

Due to the inherent risks involved, stabilised and dewatered sludge is commonly disposed of in non-beneficial ways such as by incineration or in landfill sites. However, beneficial methods of disposal are practised. One such recent innovative method has been in the making of bricks. Still the most common beneficial use of sludge is in agriculture as a soil conditioner. Adding sludge to the soil provides nitrogen and phosphorus and benefits the soil by improving moisture retention and soil structure (Korentajer, 1991). However, the inherent risks involved in using sludge in this manner limit the extent to which this option can be exploited. In conventional stabilisation and dewatering processes the sludge is not disinfected with the result that it still contains high levels of pathogens and cognizance must also be taken of the content of heavy metals and other toxic chemicals (Ekama, 1992)

1.2 SLUDGE TREATMENT AND MANAGEMENT FOR LAND DISPOSAL

1.2.1 Regulations Regarding the Usage of Sludge

In order to reduce the health risks involved with the agricultural use of sludge, many countries now have regulations or guidelines that stipulate disposal options dependent on the type of sludge treatment received and the level of toxic materials, primarily concerning heavy metals, in the sludge. The guidelines for the disposal of sewage sludge by land application in South Africa have been prepared by the Department of National Health and Population Development (DNH&PD, 1991) in terms of the Health Act (Act 63 of 1977). These guidelines are summarised in Table 1.1 below.

Table 1.1 Classification of Sewage Sludge for Beneficial Use or Disposal on Land (from DNH&PD, 1991)

SLUDGE TYPE	ORIGIN/TREATMENT (EXAMPLES)	CHARACTERISTICS/QUALITY STANDARD
TYPE A	Raw sludge Cold digested sludge Septic tank sludge Oxidation pond sludge Night Soil	• Usually unstabilised and can cause odour nuisance and fly breeding • Contains pathogenic organisms • Variable metal content
TYPE B	Anaerobic digested sludge (heated) Surplus activated sludge Humus tank sludge	• Fully or partially stabilised - should not cause significant odour nuisances or fly breeding • Contains pathogenic organisms • Variable metal and inorganic content
TYPE C	Pasteurised sludge Heat treated sludge Lime stabilised sludge ǂ Composted sludge Irradiated sludge Fumigated sludge	• Certified to comply with the following quality requirement: (If not certified this sludge is considered a Type B sludge) -Stabilised - Should not cause odour nuisance or fly breeding -Contain no viable *Ascaris* Ova per 10g dry sludge -Maximum 0 Salmonella organisms per 10g dry sludge -Maximum 1000 Faecal Coliforms per 10g dry sludge, immediately after treatment (disinfection/sterilisation) -Variable metal and inorganic content
TYPE D	Pasteurised sludge Heat treated sludge Lime stabilised sludge ǂ Composted sludge Irradiated sludge Fumigated sludge Produced for unrestricted use on land with or without addition of plant nutrients or other materials	• Certified to comply with the following quality requirement: -Stabilised - Should not cause odour nuisance or fly breeding -Contain no viable *Ascaris* Ova per 10g dry sludge -Maximum 0 Salmonella organisms per 10g dry sludge -Maximum 1000 Faecal Coliforms per 10g dry sludge, immediately after treatment (disinfection/sterilisation) -Variable metal and inorganic content • Maximum metal and inorganic content in mg/kg dry sludge Cadmium 20 Molybdenum 25 Arsenic 15 Cobalt 100 Nickel 200 Selenium 15 Chromium 1750 Lead 400 Boron 80 Copper 750 Zinc 2750 Fluoride 400 Mercury 10 • User must be informed about the moisture and NPK content • User must be warned that not more than 8 t/ha/yr (or kg/10 sq.m) dry sludge may be applied to land and that the pH of the soil should preferably be higher than 6

ǂ Composted Sludge: 55-65°C for 5 days or >65°C for 3 days

The guidelines recognise four categories of sludge depending on the degree of sludge stabilisation, pathogen disinfection, and the heavy metal and toxic element content of the sludge viz:

Type A : not stabilised or pasteurized
Type B : stabilised but not pasteurized
Type C : stabilised and pasteurized but with variable metal content
Type D : as above but with heavy metal content below specified levels

1.2.2 Influence and Effects of the Regulations on Current Treatment Practice

In South Africa, the most common method of sludge treatment in use today is anaerobic digestion. It has the attraction of being a relatively stable process, allowing for the exploitation of surplus digester gas as a usable energy source. While some degree of pathogen reduction takes place, it is generally accepted that conventional anaerobic treatment alone for stabilisation is not sufficient to also disinfect the sludge. In terms of the above guidelines anaerobically digested sludge is classified as a type B sludge and therefore unsuitable for many agricultural crops as a soil conditioner.

Certain sludge treatment systems can both stabilise and disinfect the sludge, for example, wet air oxidation (Zimpro), forced aeration composting and autothermal thermophilic aerobic digestion (ATAD). However, few can achieve this level of treatment at a low cost without significant disadvantages. The ATAD process allows for simultaneous stabilisation and disinfection of the sludge. Stabilisation is achieved by the oxidation of the biodegradable organic material in the sludge, whilst disinfection occurs thermally at thermophilic temperatures (>50°C) by the heat generated biologically through the stabilisation oxidation reactions, i.e. autothermally. The process does however require a substantial input of energy for oxygenation and mixing to achieve the thermophilic temperatures.

Whilst the ATAD process has gained some popularity in Europe, in particular in West Germany (Heideman, 1989), it would seem inappropriate to consider the process as a viable option for South Africa where considerable installed anaerobic digester capacity already exists. To convert existing anaerobic digesters to ATAD would change anaerobic digestion from an energy producing system (through methane production), to an energy consuming system (through oxygenation). A more appropriate option to consider is short residence ATAD as a pre-treatment disinfection stage prior to conventional anaerobic digestion. The combination of ATAD and anaerobic digestion in a two-stage process is known as Dual Digestion. The dual di.. stion process combines the advantage of ATAD

by providing a large degree of disinfection with that of anaerobic digestion by providing energy efficient stabilisation.

1.3 THE DUAL DIGESTION PROCESS: AN OPTION FOR THE FUTURE

1.3.1 Describing the Dual Digestion Process

The dual digestion process consists of an autothermal thermophilic aerobic first stage and a mesophilic anaerobic second stage. In the aerobic pre-treatment stage, organics present in the feed sludge are used as substrate for growth by thermophilic bacteria. The bacteria use oxygen for respiration and obtain their energy from the biochemical oxidation of the organics. The conversion to new cell material is not 100% efficient and a portion of the total energy is released as heat. By containing this biologically generated heat and not allowing it to dissipate, attainment of thermophilic temperatures become possible. Due to the higher rate of metabolism which occurs at thermophilic temperatures, these high sludge temperatures can be sustained at relatively short retention times. At these short retention times only a small portion of the organic matter in the sludge is oxidised so that only partial stabilisation is achieved in the aerobic stage.

1.3.2 The Benefits of the Dual Digestion Process

The aerobic pre-treatment in the thermophilic aerobic reactor has a number of advantages for the anaerobic digester. Aerobic pre-treatment provides a conditioning effect on the sludge resulting in a more stable operation of the anaerobic stage. This is achieved primarily by the partial dissolution of particulates which reduces the minimum required anaerobic sludge age for stabilisation. In addition ammonium ions are released into solution from the breakdown of organic nitrogen, contributing to the sludge alkalinity and providing greater pH stability to the anaerobic stage. External heating of the anaerobic stage is not necessary as the heat requirement for the mesophilic digester is provided by the hot sludge from the aerobic reactor. With the low degree of sludge stabilisation occurring during the aerobic heat treatment stage, biogas production in the anaerobic digester is not adversely affected. Because mesophilic temperatures in the digester are maintained by the hot aerobic reactor feed sludge the digester gas generated is all available for other energy requirements.

The conditioning or pre-treatment effects provided by the treatment in the aerobic reactor allows for the stabilisation in the anaerobic digestion stage to be completed at short retention times (10d), making the dual digestion process a viable system for upgrading fully loaded anaerobic digesters. This single feature may well justify implementation of dual digestion with sludge disinfection being an additional benefit.

This option is particularly appropriate to South Africa where a large number of anaerobic digestion plants exist. Where sludge disinfection is required or where existing anaerobic digesters are overloaded, conversion to Dual Digestion could meet both disinfection and increased capacity requirements.

1.3.3 Dual Digestion using Pure Oxygen

A full-scale evaluation of the dual digestion system using pure oxygen was conducted at the Milnerton Wastewater Treatment Plant. The Milnerton research dual digestion plant proved the practicability of the dual digestion process using oxygen under local conditions. Many of the advantages claimed for the system were verified by the Milnerton research and valuable information was gained in respect of process stoichiometry and biokinetics of heat generation for the aerobic reactor, sludge stabilisation and gas production in the anaerobic digester, sludge quality and dewaterability (De Villiers *et al*, 1992 and Messenger *et al*, 1992). The high cost of using pure oxygen for oxygenation in the aerobic reactor was, however, a disadvantage of the dual digestion process using pure oxygen (Laubscher *et al*, 1992) and it became logical to consider dual digester performance using air.

1.3.4 Dual Digestion using Air

Although the use of air for oxygenation in the ATAD process is now commonplace, no reports on the use of air in the dual digestion process have been received. Consequently, in 1989 the Cape Town City Council initiated a full scale research project to investigate the dual digestion process using air at the Athlone Wastewater Treatment Plant (184m^3 aerobic reactor and 1800m^3 anaerobic digester). While the research project set out initially to evaluate the performance of the dual digestion process using air alone for oxygenation of the aerobic reactor (Phase I of the investigation), from the results and conclusions drawn from the investigation the research was extended to assess the viability of using a combination of both air and pure oxygen (Phase II of the investigation). The aims and objectives of each phase in the research project are listed below.

1.4 Dual Digestion Research Objectives

1.4.1 Phase I: Oxygenation with Air

The general aim of the research programme (Phase I) was to demonstrate the practicability of the dual digestion process, employing air rather than pure oxygen to successfully disinfect and stabilise sewage sludge. Whilst the initial motivation for the installation of the dual digestion plant was based specifically on the need to provide increased digestion capacity (disinfection at that time was considered to be an added

benefit) the specific objectives of the investigation were planned to cover all the claimed benefits of the process. The objectives were as follows:

1 **Conditioning:** To assess the conditioning effects of aerobic pre-treatment on; (a) reducing the minimum required retention time for subsequent anaerobic digestion which would increase digester capacity, and (b) providing greater pH stability to the digestion stage.

2 **Disinfection:** To demonstrate that sufficiently high temperatures can be achieved in the aerobic reactor, with a typical wastewater sludge, using a simple aeration system such that (a) a satisfactory degree of disinfection is achieved and that (b) the subsequent heat requirements for optimum mesophilic anaerobic digestion are met.

3 **Stabilisation:** To assess the quality of the final sludge in terms of stability (%VS removal), fermentability (gas production), odour and dewaterability (specific resistance to filtration); and to compare these with conventional anaerobic treatment.

4 **Design:** To define the aeration and heating requirements to achieve autoheating with air oxygenation and to establish the minimum practical aerobic and anaerobic retention times for the process. To develop a mathematical model to enable prediction of reactor temperatures and to recommend criteria for the design and operation of future plants.

5 **Operation:** To estimate the practical operation problems of a full scale system over an extended period of operation and to make an assessment of the reliability of the process.

6 **Costs:** To estimate the capital, operation and maintenance costs for the system and to make comparisons with conventional anaerobic digestion and the dual digestion process using oxygen.

The results and conclusions drawn from Phase I of the Athlone dual digestion project, in terms of the above objectives, are presented in Chapter 4. The time and duration of Phases I and II is shown in Figure 1.1 below.

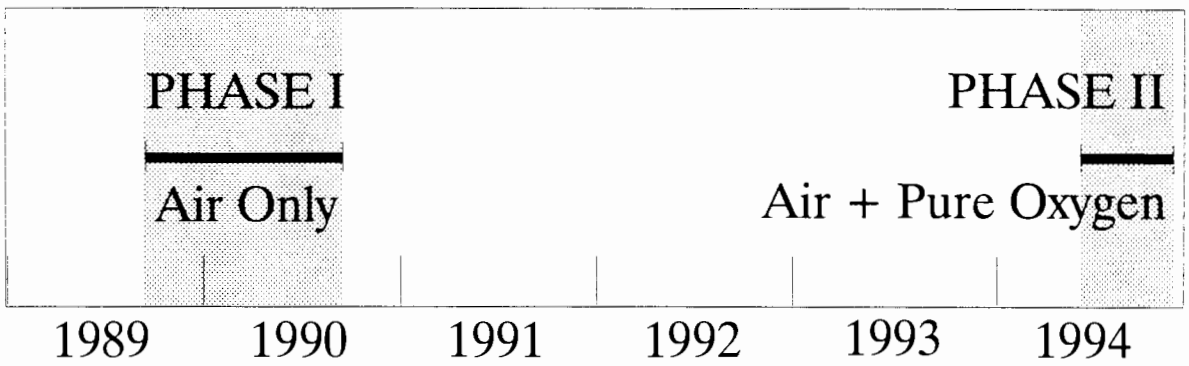


Figure 1.1 Schematic Showing the Time Frame and Duration of Phases I and II of the Athlone Dual Digestion Project

1.4.2 Phase II: Oxygenation with Air and Pure Oxygen

In 1994 the Cape Town City Council commenced with Phase II of the research project, with the general aim of demonstrating the practicability of the dual digestion process employing a combination of both air and pure oxygen to successfully disinfect and stabilise sewage sludge. It was considered that by supplementing the air process with pure oxygen, increased treatment capacity and greater process control of the aerobic reactor could be achieved. Further, the benefits of the pure oxygen process would be retained at lower operating costs with a significant proportion of the oxygen demand provided by the transfer of atmospheric oxygen from the air stream. The specific objectives of the investigation were as follows:

1 **Oxygenation Characteristics:** To determine the pure oxygen supply rate(s) required for a range of different operating conditions, viz:

Sludge feed volume	reactor temp	ambient temp
120-180 m^3/d	50-60°C	15-25°C

and to determine the oxygen transfer efficiencies which can be effected by the pure oxygen injection equipment at the different supply rates.

2 **Foaming:** To assess the influence which foam formation has on reducing the required pure oxygen supply rate(s). To establish the optimum conditions for foam formation and to determine the requirements for operation with a stable foam layer. The dependence of foam formation on influent air flow rate, reactor temperature and feeds solids concentrations will be examined at laboratory scale.

3 **Digester Heating Requirements:** To establish the seasonal heating requirements for optimum operation of anaerobic digester, i.e. the desired aerobic reactor temperature and retention time which will enable the anaerobic digester to operate at 36-39°C. This will entail conducting a steady state heat balance on the anaerobic digestion stage of the Athlone dual digester.

4 **Costs:** To determine the capital, operation and maintenance costs associated with pure oxygen injection and to make comparison with conventional anaerobic digestion and the dual digestion process using pure oxygen alone.

5 **On Site Generation:** To assess the viability of employing a (vacuum swing absorption) VSA plant for generating oxygen on site. This would include the sizing of an appropriate VSA plant.

6 **External Heating:** To examine the feasibility of pre-heating the feed sludge, and then if possible, determine the reduction in oxygen supply rate which can be effected by pre-heating.

The results and conclusions drawn from Phase II of the Athlone dual digestion project, in terms of the above objectives, are presented in Chapter 6.

1.5 Scope and Layout of Thesis

Chapter 1 starts with a general description of municipal wastewater treatment, followed by a discussion on the treatment and disposal of sewage sludge. The recently published guidelines (from DNH&PD, 1991), which regulate the use of sewage sludge for disposal on land, are presented. The need to pasteurise sewage sludge prior to any form of land application for agricultural purposes is discussed. The benefits and suitability of the dual digestion process in providing both pasteurisation and stabilisation are presented. The relatively high operating cost of the pure oxygen dual digestion process is raised, a factor which formed part of the motivation for undertaking this research. The aims and objectives of phase I of the investigation, where oxygenation is with air alone, are listed. Mention is made of the motivation for phase II of the investigation, where oxygenation is with air and pure oxygen, followed by a complete listing of the aims and objectives for phase II.

Chapter 2 presents an overview of the Athlone Wastewater treatment plant to give an impression of the type and operation of the plant and therefore of the quality and

quantity of the sewage sludge produced. This is followed by a description of the layout of the full scale dual digestion plant at Athlone. The operation and monitoring of the plant during both phases of the investigation are then discussed. The Chapter concludes with a description of the laboratory scale anaerobic digesters, which were operated to test the conditioning effects of aerobic pre-treatment, and the pilot scale aerobic reactor which was operated in order to establish some of the basic characteristics of foam formation.

Chapter 3 starts with a review of the biokinetics and stoichiometry of biological heat generation. This is followed by the development of the equations from which the biological heat generation and oxygen utilisation rates could be calculated from the measurable parameters, namely sludge, air and pure oxygen flow rates, vent gas oxygen concentration and the assorted temperatures across the system.

Chapter 4 presents the results obtained from phase I of the investigation, where oxygenation of the aerobic reactor is with air alone. A summary of overall plant performance is first presented, followed by a more detailed evaluation of aerobic reactor performance dealing specifically with the characteristics of the aeration system and the biological heating rate. Detailed aspects of overall system performance are then discussed which includes, the relationship between biological heating, oxygen utilisation and volatile solids destruction, the conditioning effects of aerobic treatment, and the quality of the final sludge. The Chapter concludes with the results of the laboratory scale study which evaluated the effect of aerobic conditioning on anaerobic digester stability.

Chapter 5 presents simple mathematical models for the steady state design of air oxygenated aerobic reactors in dual digestion systems, with allowance made for both foaming and non-foaming conditions. The models are then used to predict the effects of incorporating pure oxygen supplementation and feed sludge pre-heating on the performance of the aerobic reactor. The model for predicting pure oxygen supplementation rates is used in the design and operation of the pure oxygen injection system during phase II of the investigation.

Chapter 6 presents the conclusions drawn from phase I of the investigation. This includes an overall assessment of phase I of the investigation, followed by the specific conclusions drawn in relation to the original laid down objectives. The viability of the dual digestion process using air is discussed, followed by the recommendation for phase II of the investigation, namely to assess the viability of pure oxygen supplementation. The Chapter concludes with a listing of the proposed aims and objectives for phase II.

Chapter 7 presents the results obtained from phase II of the investigation, where oxygenation of the aerobic reactor is with air and pure oxygen. A summary of overall plant performance is first presented, followed by a more detailed evaluation of aerobic reactor performance dealing specifically with the characteristics of the pure oxygen injection system and the biological heating rate. Further discussion is then presented on the relationship between biological heating, oxygen utilisation and volatile solids destruction. With the anaerobic digester operating at thermophilic temperatures a full discussion is presented on the quality of the final sludge.

Chapter 8 presents a general mathematical model for the steady state design of the dual digestion process. The model is able to predict required oxygen utilisation rates in the aerobic reactor when oxygenation is with air and/or pure oxygen, over a wide range of different operating conditions. The approximate retention times at which the aerobic reactor becomes substrate limited is predicted for different feed sludge types. The model estimates volatile solids destruction across both stages in the process together with biogas production from the digester. The effects of (1) employing biogas to pre-heat the feed sludge and (2) incorporating sludge/sludge heat exchangers on reducing the required oxygen utilisation rates are examined. The minimum digester retention time to ensure satisfactory sludge stabilisation is predicted. For comparison purposes, the model is capable of simulating conventional mesophilic and thermophilic anaerobic digestion.

Chapter 9 presents the conclusions drawn from phase II of the investigation. This includes an overall assessment of phase II of the investigation, followed by the specific conclusions drawn in relation to the original laid down objectives. The viability of the dual digestion process using air with pure oxygen supplementation is discussed, followed by the recommendations for future operation of the dual digestion plant at Athlone. The Chapter concludes with an assessment of the overall success of the investigation and summarises the main findings of the report.

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CHAPTER 2

OPERATION AND MONITORING

2.1 THE ATHLONE WASTEWATER TREATMENT PLANT

2.1.1 Location and Catchment Area

The Wastewater Treatment Plant at Athlone, Cape Town employs trickling filters and an activated sludge process to treat an average dry weather flow (ADWF) of 120 Ml/d. The location of the catchment area for the Athlone Plant is shown in Figure 2.1. The area is 82.6 km^2 in size and serves a population of approximately 360,000 people.

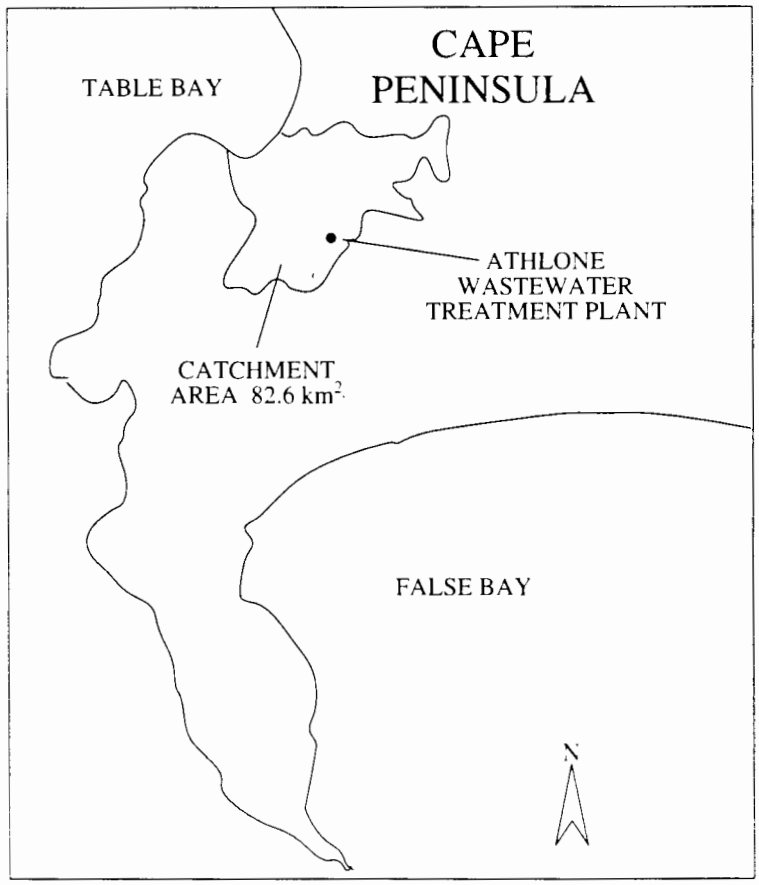

Figure 2.1 Location and Catchment Area Plan: Athlone Wastewater Treatment Plant

The raw wastewater contains a significant fraction of industrial effluent and infiltration of groundwater is noticeable during periods of heavy rainfall in winter. The flow rate and chemical values for the main parameters characterising the incoming wastewater during phase I of the dual digestion investigation (1989/90) are detailed in Table 2.1.

Table 2.1 Raw Wastewater Characteristics: Athlone Wastewater Treatment Plant 1989/90 During Phase I of the Investigation

Parameter		mean	min	max
Flow Rate	Ml/d	105	35	198
Suspended Solids	mg/l	330	150	870
Chemical Oxygen Demand	mg(O)/l	860	430	2400
Free and Saline Ammonia	mg(N)/l	24	9.1	34
Organic Nitrogen	mg(N)/l	23	7.5	94
Total Phosphorus	mg(P)/l	9.9	5.9	22
pH	-	7.3	5.5	7.8
Conductivity	Ms/m	120	87	140
Alkalinity	mg($CaCO_3$)/l	220	95	310

2.1.2 The Historical Development of the Athlone Plant

The first Athlone Wastewater Treatment Plant was constructed in 1921, and consisted of 4 Imhoff tanks, brushwood filters and two-storey humus tanks. The main intercepting sewers gravitated to a strategic locality on the Black River at Raapenberg where a major pumping station pumped the sewage to the Plant.

To cope with the rapid growth in urban development, a new treatment plant was constructed in 1939. Experimental work on the activated sludge process was not encouraging, and so conventional methods employing primary sedimentation and stone bio-filtration were employed to treat the wastewater. Further additions increasing treatment capacity to 36 Ml/d were effected during 1952 and 1953 and it was at this time that the first of the existing anaerobic digesters were constructed to treat the primary/humus sludge mix produced by the sedimentation/bio-filtration process. Duplication of the Plant, initiated in 1956 to cater for an additional flow of 44 Ml/d, was completed in 1961 and incorporated additional mechanical equipment, bio-filtration treatment units and anaerobic digesters. Also, sludge gravity thickeners and maturation ponds were added, the latter to meet the requirements of the Athlone Power Station, which was reusing the treated effluent as cooling water.

The implementation of more stringent standards for the discharge of effluent in 1962, arising from the Water Act 54 of 1956, meant that the Athlone Plant which had been designed and completed to the British Royal Commission Standards of 1903, no longer complied with legal requirements. A major programme for modernisation of the plant was therefore initiated in 1973.

The modernisation and extension of the then 80 Ml/d Athlone Plant to 120 Ml/d was implemented in 1980. The design included flow balancing, fat removal and an activated sludge process for the treatment of bio-filter effluent. Finally sand filters were installed to ensure complete removal of the suspended matter. The construction and installation of additional sludge handling facilities is the only outstanding item in the modernisation plan, and it was in this respect that the dual digestion investigation was initiated in 1989. The claimed benefit of the dual digestion process in reducing the required anaerobic digestion time was seen as a possible solution for increasing digestion capacity at the Plant.[1]

2.1.3 Current Treatment Practice at the Athlone Plant

A schematic diagram of the wastewater and sludge treatment processes currently employed at the Athlone Plant (1990) is shown in Figure 2.2.

Wastewater Treatment

After screening and grit removal, fats and grease present in the sewage are removed by dissolved air flotation. The flow is then equalised by a set of balancing tanks. The equalised flow passes through a set of primary sedimentation tanks (PST) for primary clarification. A portion of the PST overflow, which can be varied but generally is about 50%, is distributed to the bio-filtration plant consisting of bio-filters and humus tanks. The remainder of the PST overflow, which by-passes the filters, is mixed with the effluent from the humus tanks. This mixture is then fed to the activated sludge plant.

The activated sludge plant is operated in the modified Lutzack-Ettinger mode for COD removal, nitrification and denitrification. The mode of operation is flexible allowing the operator to optimise nitrogen removal. Secondary settling tanks provide secondary clarification. A portion of this effluent (about 20%) is clarified further by sand filtration prior to acceptance by the Athlone power station. The remainder passes through a series

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[1] The motivation for the installation of the dual digestion plant at Athlone was based specifically on the need to provide increased digestion capacity. Disinfection if achieved was considered at that time to be an added benefit and therefore not a primary objective.

of maturation ponds prior to discharge to the Vygekraal river, which flows into Table Bay via the Black River.

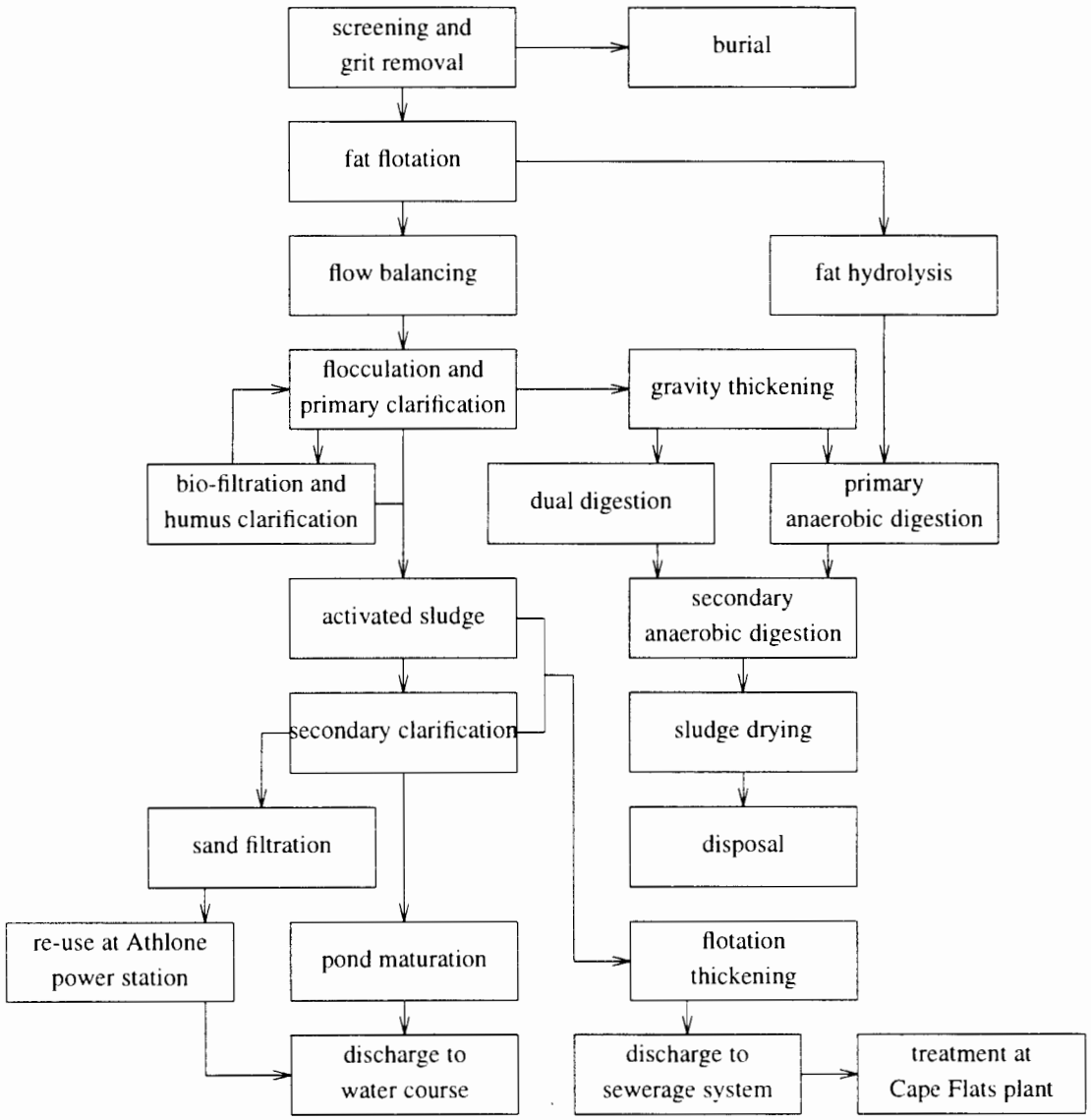

Figure 2.2 Schematic of the Athlone Wastewater Treatment Plant

Sludge Treatment

After gravity thickening the primary sludge is stabilised in conventional mesophilic anaerobic digesters and dewatered on drying beds. The fats removed by flotation are hydrolysed in an acid digester. The resulting liquor, containing large quantities of acetic acid is pumped to the activated sludge plant. The inert solids from the acid digester are pumped to the anaerobic digestion plant. The humus sludge from the bio-filtration plant is recycled up-stream of the primary settling tanks, where after flocculation with the settleable solids in the wastewater is removed with the primary sludge in the PST's. The waste activated sludge is thickened by dissolved air flotation and the resulting float is discharged to the sewerage system for treatment at the Cape Flats Wastewater Treatment

Plant. Consequently at Athlone, only the primary/humus sludge is anaerobically digested and it was this sludge that was treated in the dual digestion plant.

2.2 THE ATHLONE DUAL DIGESTION PLANT

2.2.1 The Dual Digester

The dual digestion plant was constructed by converting a decommissioned primary anaerobic digester (see Figure 2.2). A reinforced concrete aerobic reactor was constructed centrally within the structure of the anaerobic digester (see Figure 2.3). The objective of this type of design was to keep wall heat losses to a minimum and to simplify the manner in which sludge is transferred between the aerobic reactor and anaerobic digester.

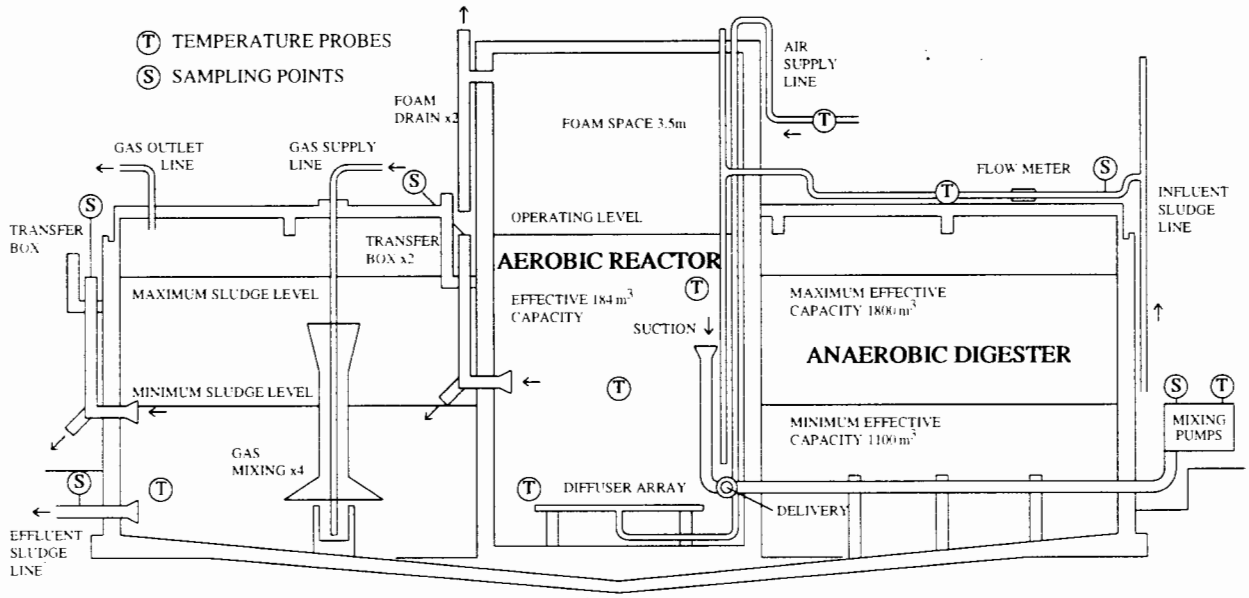

Figure 2.3 Schematic of the Dual Digestion Plant at Athlone

2.2.2 The Aerobic Reactor

The aerobic reactor (inner core) has an effective liquid operating capacity of 184 m^3 and additional capacity is provided to accommodate a foam layer of up to 3m in depth. Two foam drains, located just below the roof of the reactor, direct any foam spillage into two sludge transfer boxes which transferred sludge from the aerobic reactor to the anaerobic digester. Air was supplied by course bubble aeration and mixing was effected by a recirculation pump (see Section 2.3.4 and 2.3.6 below for details). During phase II, pure oxygen was injected into the sludge recirculation line, which had been modified for this purpose, to supplement the coarse bubble aeration system (see Section 2.3.5 and 2.3.7 below for details).

2.2.3 The Anaerobic Digester

The anaerobic digester has an operating capacity ranging between 1100 and 1800 m^3. Hot sludge from the aerobic reactor is transferred to the anaerobic digester via two transfer boxes; the sludge is transferred by displacement each time feed sludge is pumped into the aerobic reactor. When the digester is operated at full capacity, digested sludge is transferred to the secondary digester via a single transfer box. When the digester is operated at minimum capacity, digested sludge is withdrawn from the base of the digester and pumped to the secondary digester. When the transfer box is not in use a float indicator is used to control the sludge level. Mixing of the contents of the digester is provided by gas recirculation. This gas recirculation system connected all the primary digesters to the gas holder, with the result that it was not possible to measure the gas production, composition, and vent gas heat loss from an individual digester i.e. the anaerobic digester of the dual digestion system (see Section 2.3.10).

2.3 Operation of the Athlone Dual Digestion Plant

2.3.1 Feed Sludge Thickening

The feed sludge for the dual digestion system was drawn from two primary sludge gravity thickeners which received sludge from twelve primary settling tanks. The primary sludge is derived from the readily settleable components in the raw wastewater, after screening and grit removal, and consists mostly of organic material (about 81% volatile). The flow, total solid and volatile solid content of the primary sludge produced at the Athlone Wastewater Treatment Plant during the phase I evaluation period are given in Table 2.2.

Table 2.2 Primary Sludge Data During Phase I: Athlone Wastewater Treatment Plant 1989/90

Parameter		mean	min	max
Flow Rate	m^3/d	1000	-	-
Total Solids	kg/m^3	25.0	4.6	55.6
Volatile Solids (dry mass basis)	%	80.8	69.9	87.9
pH	-	5.8	4.7	6.6

The primary settling tanks were de-sludged in rotation once every hour for a period of approximately 10 minutes each to provide a semi-continuous feed to the gravity thickeners. The gravity thickeners were de-sludged together once every hour for a period of approximately 15 minutes. The overflow (supernatant) from the gravity thickeners was returned upstream of the primary settling tanks. The frequency and duration of de-sludging of the primary settling tanks and gravity thickeners is estimated by the operating staff on the basis of a visual assessment of the thickness (concentration) of sludge in the PST or gravity thickeners.

Prior to the commissioning of the dual digestion system, the gravity thickened sludge was pumped to operating primary sludge anaerobic digesters. During the dual digestion evaluation period, the aerobic reactor feed sludge was pumped from this flow of gravity thickened sludge.

2.3.2 Control of Aerobic Reactor Retention Time

During phase I of the investigation, feeding of the aerobic reactor took place at either 2 or 4 hourly intervals and the duration of the pumping periods depended on the specified aerobic reactor retention time. Two piston pumps were available to pump sludge to the aerobic reactor. With both pumps running the combined discharge rate was approximately 0.50 m^3/min. The duration of the feeding periods for different retention times ranging from 3 to 8 days is listed in Table 2.3 below (the system was fed semi-continuously and not on a draw and fill basis because pasteurisation was not a priority). The quantity of sludge pumped to the aerobic reactor was monitored by an in-line flow meter (see Section 2.4.4). The feed sludge was sampled during each feeding session to produce a daily composite sample for analysis (see Section 2.4.5).

Table 2.1 Schedule and Duration of Feeding Times for the Aerobic Reactor During Phase I

Hydraulic Retention Time	d	7.7	5.1	3.8	3.1
Daily volume of sludge pumped	m^3/d	24	36	48	60
Time between batches	h	4	4	2	2
N° of batches per day	-	6	6	12	12
Volume of sludge pumped per batch	m^3	4	6	4	5
Time taken to pump each batch	min	8	12	8	10

During phase II of the investigation, much higher sludge loading rates were possible due to the injection of pure oxygen to supplement the aeration system. During this phase, feeding took place at either hourly or two hourly intervals. The duration of the feeding periods for different retention times ranging between 1 to 2 days is listed in Table 2.4 below.

Table 2.2 Schedule and Duration of Feeding Times for the Aerobic Reactor During Phase II

Hydraulic Retention Time	d	1.0	1.25	1.5	2.0
Daily volume of sludge pumped	m^3/d	184	147	123	92
Time between batches	h	1	1	2	2
N° of batches per day	-	24	24	12	12
Volume of sludge pumped per batch	m^3	7.7	6.1	10.2	7.7
Time taken to pump each batch	min	15	12	20	15

2.3.3 Transfer of Aerobic Sludge to the Anaerobic Digester

Hot aerobic reactor sludge was transferred to the anaerobic digester by displacement[2] caused by the addition of feed sludge to the reactor. The sludge transfer takes place via the two transfer boxes. The positioning of theses boxes (at liquid level) relative to the influent feed discharge point (base of the reactor) together with the mixing pattern (see Section 2.3.5) was such that short circuiting was kept to a minimum in the aerobic reactor. In order to obtain more representative samples from the aerobic reactor, samples were taken from a point on the recirculation line close to the mixing pumps (see Section 2.4.5) and not from the transfer boxes to the anaerobic digester.

[2] Because upgrading of the anaerobic digesters rather than pasteurisation of sludge was the primary objective of the project, batch feeding of the aerobic reactor was not catered for in the prototype dual digestion plant.

2.3.4 The Aerobic Reactor Aeration System

Two liquid ring compressors were available to supply air to the aerobic reactor. The air was passed into the reactor through coarse bubble diffusers set in the base of the reactor. The operation of the liquid ring compressors were such that it could be assumed with reasonable accuracy that the delivery air flow was saturated with water vapour (see Section 2.4.11). The delivery air temperature was measured. Under normal conditions only one liquid ring compressor was operated, capable of providing a maximum possible air flow rate of 780 m^3(STP)/h. The second liquid ring compressor was available to provide back-up in the event of mechanical failure of the first. The second compressor was used to provide gas mixing in the anaerobic digester. During the commissioning of the aerobic reactor a non-steady state aeration test with ambient temperature tap water (17°C) was undertaken. The standard oxygen transfer rate (the Oxygenation Capacity *OC*) obtained was 30.5 kg(O_2)/h giving a volume specific rate of 0.166 kg(O_2)/m^3.h. The standard oxygen transfer efficiency was calculated at 14.0%. Details of the test are given in Appendix 4. On one occasion (steady state period 5) the performance of the aerobic reactor was assessed with both compressors operating simultaneously providing a combined air flow of 1200 m^3(STP)/h. No mixing in the anaerobic digester took place during this period.

2.3.5 The Pure Oxygen Injection System

Pure oxygen was injected into the aerobic reactor sludge recirculation line during phase II of the investigation (1994). The injection point was immediately downstream of the mixing pumps. The injection device was a venturi type injector (Vitox patented by AFROX) of similar design to that used on the pure oxygen aerobic reactor at Milnerton (see Messenger *et al*, 1991 for details). Liquid oxygen was stored on-site in a cryogenic storage vessel 6000kg in capacity. After passage through an evaporator the flow of oxygen gas, which could be regulated by adjusting an appropriate valve, was recorded by a rotameter flow meter, before passing to the venturi injector.

2.3.6 The Aerobic Reactor Mixing System: Phase I

During phase I (1989/90), mixing of the aerobic reactor contents was provided by two 7.5 kW centrifugal mixing pumps with a pumping design capacity of 396 m^3/h each. Each pump was operated alternately at fixed time intervals (usually 4hrs) to provide continual mixing and to distribute the workload between each pump. The discharge pipe for the mixing recirculation flow was 400 mm ϕ giving a calculated 0.9 m/s flow velocity entering the aerobic reactor. This flow was discharged tangentially into the reactor producing a horizontal circular mixing pattern. The mixing effect of the aeration system provided vertical mixing of the sludge. The combination of the two mixing patterns produced a corkscrew pattern. During the evaluation period the power drawn by each

mixing pump (rated 7.5kW) was in excess of 10kW due to the negligible difference in head between the inlet and outlet pipes of the pumps. The mixing pumps managed to operate throughout the phase I evaluation period but shortly after its termination the electrical motors burnt out. Consequently, between phases I and II no mixing of the aerobic reactor, via the recirculation pumps, took place. The operation of the mixing pumps is discussed further in Section 3.9.

2.3.7 The Aerobic Reactor Mixing System: Phase II

In order to try and satisfy the requirements for pure oxygen injection[3], the aerobic reactor mixing system was modified for operation during phase II. In order to obtain the necessary pressure in the recirculation line and the required velocity at discharge, the original 400mmϕ pipework downstream of the pumps was replaced with a 200mmϕ pipe. This pipe, instead of passing directly back into the reactor, passed 90° around the outside of the digester, and was connected to a 137mmϕ pipe which carried the sludge through the digester and into the aerobic reactor. The additional length of recirculation line (approximately 15m) was considered important because it increased the contact time between sludge and oxygen prior to the pressure being reduced to hydrostatic pressure inside the aerobic reactor.

At the start of phase II (days 1-25), one pump was operated fitted with a 55kW motor. Between days 26-40, two pumps were operated in parallel, with the second pump fitted with a 75kW motor. As neither pumping condition produced the desired conditions for efficient oxygen injection, the pipework was modified to allow the two pumps to operate in series (days 74-152). At this time, the 75kW motor was fitted with a variable speed drive to allow for operation at different rotational speeds.

Full details of the modifications which were effected to the aerobic reactor mixing system prior to and during phase II are discussed in Appendix 10 below. The effect of recirculation line flow rate, pressure, and discharge velocity on the pure oxygen transfer efficiency is discussed in Section 7.3.3 below.

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[3] In order to achieve the necessary pure oxygen injection rates at a transfer efficiency in excess of 80%, the pressure in the sludge recirculation line needed to be increased to above 300kPa, the flow rate in the recirculation line was required to be in excess of 700m^3/h, and the velocity at the point of discharge needed to exceed 15m/s.

2.3.8 Foam Control in the Aerobic Reactor

Large quantities of foam were produced on frequent occasions during the phase I of the investigation, often for extended periods of time. While large spillage of foam are not acceptable for obvious practical reasons, the benefits provided by the presence of a large foam layer made it desirable to manage the foam rather than try and eliminate it. According to Wolinski (1985) the performance of the aerobic reactor is improved by the presence of a significant foam layer through increased oxygen transfer, enhanced biological activity and improved heat insulation. Consequently in the design of the aerobic reactor, sufficient head space to accommodate a foam layer of up to 3.5m in depth was allowed for. Under foaming conditions with an air flow rate of 760 m^3(STP)/h it became evident that this 3.5m headspace was insufficient to prevent spillage. Foam control was achieved by reducing the air flow rate until the foam level stabilised at 3m in depth, a control strategy which benefited the process by reducing vent gas vapour heat losses. To deal with excessive build up of foam, two overflow channels were provided to allow the foam to pass, via the sludge transfer boxes (see Figure 2.3), to the anaerobic digester.

2.3.9 Anaerobic Digester Operating Level Control

Treated sludge from the anaerobic digester was transferred to a secondary digester by displacement via one transfer box located at the outer wall of the anaerobic digester (see Figure 2.3). The final sludge from the system was sampled at this point (see Section 2.4.5), by compositing over 24 hours a grab sample taken at 8h intervals during the aerobic reactor feeding period.

At the start of the evaluation period in phase I the digester was operated at the minimum operating capacity 1100 m^3, however this caused practical problems in that, on occasion sludge was syphoned back into the digester from the secondary digester. Also the efficiency of gas mixing was significantly reduced at low liquid levels in the digester (<1500m^3). Consequently, the digester was operated at the maximum operating capacity of 1800m^3 throughout the investigation to ensure good mixing. This made it difficult to meet one of the major objectives for phase I of the investigation (see Section 1.6) i.e. to establish the minimum anaerobic digestion time for stable treatment. It is claimed that the sludge is conditioned in the thermophilic aerobic reactor such that the minimum required anaerobic digestion retention time is reduced (see Section 3.1.9). Because of its importance in the evaluation of the dual digestion process, this objective was tested at laboratory scale during phase I (see Section 2.5) in a similar investigation to that of Izett *et al* (1992).

During phase II, because much higher loading rates on the aerobic reactor were possible due to pure oxygen injection, the retention time in the anaerobic digester could be lowered to below 10 days. In addition, because of the high sensible heat content of the aerobic sludge, and no deliberate temperature control in the anaerobic digester, the anaerobic digester which operated in the mesophilic range during phase I increased into the thermophilic range during phase II. Consequently, under such conditions, it was possible to fully evaluate the quality of the final sludge and assess the effects of aerobic conditioning.

2.3.10 Anaerobic Digester Mixing

Mixing of the anaerobic digester contents was effected by recycling biogas, using one of the liquid ring compressors, through 4 draft tubes, equally spaced around the inside of the digester. The biogas generated by the anaerobic digester became part of the gas in the communal gas collection, storage, and recirculation network of the primary sludge anaerobic digesters of which the dual digester was still a part. Consequently the gas production in the anaerobic stage of the dual digester could not be directly measured. Also no gas meters were installed to monitor the volumes of biogas entering (recirculated) and leaving the digester because large fluctuations in gas pressure made estimation of the gas flow impractical. As a result of these factors it was not possible to determine both the gas composition and production and the vent gas water vapour heat losses from the anaerobic digester.

2.4 MONITORING OF THE ATHLONE DUAL DIGESTION PLANT

2.4.1 The Monitoring Parameters

The parameters which were monitored during both phase I and phase II are summarised below; Table 2.4 presents the parameters monitored to determine the aeration characteristics and performance of the aerobic reactor and Table 2.5 presents the parameters monitored to determine the system sludge characteristics and anaerobic digester performance. The frequency at which the analysis was performed is also indicated.

Table 2.4 Monitoring Parameters Employed in Evaluating the Aeration Characteristics of the Aerobic Reactor, During both Phase I (Oxygenation with Air Only) and Phase II (Oxygenation with Air and Pure Oxygen)

Parameter	Symbol	Units	Frequency
Dry Air Flow Rate	$Q(AIR)_{in}$	m^3(STP)/h	daily
Pure Oxygen Mass Flow Rate	$M(O_2)_{in}^{O2}$	kg(O_2)/h	daily
Reactor Gas Temperatures	$T(AIR)_{in}$, $T(AIR)_{out}$	°C	daily
Vent Gas Analysis Performed During Phase I			
%O_2 Concentration in Vent Gas	$\%(O_2)_{out}$	%v/v	†
%CO_2 Concentration in Vent Gas	$\%(CO_2)_{out}$	%v/v	†
Vent Gas Analysis Performed During Phase II			
%O_2 Concentration in Vent Gas with Pure Oxygen Supplementation **On**	$\%(O_2)_{out}^{AIR+O2}$	%v/v	†
%O_2 Concentration in Vent Gas with Pure Oxygen Supplementation **Off**	$\%(O_2)_{out}^{AIR}$	%v/v	†

Notes: 1) Subscripts *in* and *out* denote influent and effluent gas values for the aerobic reactor.
2) Superscripts *AIR+O2* and *AIR* denote oxygenation with either *air and pure oxygen* or *air alone* whilst monitoring the vent gas during phase II.
† Performed intermittently (est. 2 to 3 times per week) during steady state periods only.

2.4.2 Process Reliability

One of the requirements for both phases of the investigation was to gauge the reliability of the process in terms of ease of operation and to note any mechanical or practical difficulties. To assess the reliability of the process in these terms the operating staff kept daily records in which any faults or problems were noted. A discussion on this aspect is given in Section 4.2.2.

Table 2.5 Monitoring Parameters Employed in Evaluating the Sludge Characteristics During both Phase I (Oxygenation with Air Only) and Phase II (Oxygenation with Air and Pure Oxygen)

Parameter	Symbol	Units	Frequency
Feed Sludge Flow Rate	$Q(SL)_{in}$	m^3/d	daily
Sludge Temperatures	$T(SL)_{in}$, $T(SL)_r$, $T(SL)_d$	°C	daily
Total Solids Concentration	TS_{in}, TS_r, TS_d	$kg(TS)/m^3$	daily
Volatile Solids Concentration	VS_{in}, VS_r, VS_d	$kg(VS)/m^3$	daily
Chemical Oxygen Demand	COD_{in}, COD_r, COD_d	$kg(O)/m^3$	daily
Conductivity	$COND_{in}$, $COND_r$, $COND_d$	Ms/m	daily
Ammoniacal Nitrogen	$NH4_{in}$, $NH4_r$, $NH4_d$	$g(N)/m^3$	daily
pH	pH_{in}, pH_r, pH_d	-	daily
Volatile Acid Alkalinity	VA_{in}, VA_r, VA_d	$mg(CaCO_3)/l$	daily
Bicarbonate Alkalinity	BA_{in}, BA_r, BA_d	$mg(CaCO_3)/l$	daily
Faecal Coliforms	FC_{in}, FC_r, FC_d	/cl	†
Total *Ascaris* Ova	ASC_{in}, ASC_r, ASC_d	/g	†
Viable *Ascaris* Ova	VI_{in}, VI_r, VI_d	%	†
Specific Resistance to Filtration ‡	SRF_d	m/kg	†

Notes: 1) Tests on sludge were on daily composite samples.
2) Subscripts *in,r* and *d* denote influent (feed), reactor and digester values respectively.
† Performed intermittently (est. 1 to 2 times per week) throughout the evaluation period.
‡ The SRF test was only performed on the final sludge from the anaerobic digester.

2.4.3 Temperature Measurement

Seven PT100 resistance thermometers were positioned around the aerobic reactor to measure temperatures. The positions of the thermometers are shown in Figure 2.3 and were as follows:

- 3 within the aerobic reactor
- 2 in the aerobic reactor recirculation loop $T(SL)_r$
- 1 in the feed sludge line to the aerobic reactor $T(SL)_{in}$
- 1 in the air delivery line to the aerobic reactor $T(AIR)_{in}$
- 1 in the anaerobic digester $T(SL)_d$

The temperatures from the thermometers placed in the recirculation loop were recorded by chart recorder. Gauges were used to measure temperatures from the other thermometers with the operating staff manually recording temperatures every half hour

on specially designed log sheets. Calibration of the temperature probes were checked at monthly intervals.

The effluent (vent) gas stream temperature $T(AIR)_{out}$ from the aerobic reactor was measured by a combustion analyser which also measured the oxygen and carbon dioxide fractions in the vent gas stream. The accuracy of the results obtained by the combustion analyser were checked regularly by doing parallel ORSAT tests (see Messenger *et al*, 1992 for details).

2.4.4 Influent Sludge Flow Rate

An in-line flow meter, placed close to the point of entry to the aerobic reactor, measured the quantity of sludge pumped into the reactor. By controlling the daily sludge volume pumped to the aerobic reactor the desired aerobic reactor retention time was maintained. The relationship between the daily volumetric sludge flow rate to the reactor and the hydraulic retention time is given by Eq 2.1:

$$R_h = \frac{V_p}{Q(SL)_{in}} \quad \text{...days} \quad (2.1)$$

where:

R_h = Aerobic reactor hydraulic retention time

V_p = Aerobic reactor process volume

$Q(SL)_{in}$ = Volumetric flow rate of feed sludge to the aerobic reactor (m^3/h)

The vaporisation of water from the sludge liquid to the reactor vent gas stream can reduce the sludge volume leaving the reactor by up to 5 to 8 percent depending on the retention time (see Section 3.5.7). While this extends the retention time in the reactor, this effect was ignored in both the calculation of the retention time with Eq 2.1 and the calculation of the percentage VS, TS, and COD removals.

The feed sludge flow meter was routinely checked with values obtained from the pump meter at the gravity thickeners and by volumetric comparison in the thickener tanks (displacement). Visual checking in the gravity thickeners of the volume of sludge pumped to the dual digestion system was possible because the dual digester and each primary anaerobic digester were fed in a fixed sequence and at separate times.

2.4.5 System Sludge Stream Sampling

To take account of the variation of solids concentration in the gravity thickened sludge stream, the plant operators collected a grab sample midway through each feeding session and composited these over 24 hours to make a daily composite sample.

Sludge from the aerobic reactor and anaerobic digester were sampled once every eight hours and composited over 24 hours. The aerobic reactor sludge was sampled at the recirculation mixing pump. The anaerobic digester sludge was sampled at the digester effluent transfer box when sludge was being transferred to the secondary digester.

Samples for bacteriological analyses and dewaterability studies were taken as grab samples from the aerobic reactor or anaerobic digester as required on the specific days these analyses were performed.

2.4.6 The Oxygen Supply Rate (Air Only)

The volumetric influent dry air flow rate to the aerobic reactor $Q(AIR)_{in}$ was monitored by an orifice plate flow meter with the flow recorded on a chart recorder housed in a control room. Calibration of the meter was carried out at monthly intervals. The flow rate was automatically corrected to STP, 20°C and 760 mmHg as dry air, at which temperature and pressure the density of dry air, computed from the Ideal Gas Equation, is $\rho(AIR)$ = 1.205 kg/m^3. It therefore follows that the mass flow rate of air entering the reactor is given by:

$$M(AIR)_{in} = 1.205 \times Q(AIR)_{in} \quad \text{...kg/h} \quad (2.2)$$

The percentage oxygen by mass in atmospheric air is 23.0%. It therefore follows that the mass flow rate of oxygen into the reactor is:

$$M(O_2)_{in}^{AIR} = 0.230 \times 1.205 \times Q(AIR)_{in} = 0.277 \times Q(AIR)_{in} \quad \text{...kg/(O}_2\text{)/h} \quad (2.3)$$

The oxygen supply rate ***OSR*** defines the mass of oxygen supplied per unit process volume. The process volume V_p of the aerobic reactor = 184 m^3. It therefore follows that the oxygen supply rate attributable to the transfer of oxygen from the air stream is:

$$OSR = \frac{0.277}{184} Q(AIR)_{in} = 0.00151\ Q(AIR)_{in} \quad \text{... kg(O}_2\text{)/h.m}^3 \quad (2.4)$$

Where $Q(AIR)_{in}$ is in m^3(STP)/h. In Section 3.2.9 another function for ***OSR*** is derived based on the volumetric fraction of oxygen in the vent gas (see Eq 3.37). Both approaches give the same result.

For the purposes of calculating the oxygen utilisation rate ***OUR*** (see Section 3.2) an accurate estimation of the dry effluent molar gas flow rate was made by conducting a gas component mass balance across the reactor. i.e the dry influent and effluent molar flow rates were <u>not</u> assumed equal. Whilst the assumption that the dry influent and effluent molar flow rates are <u>not</u> equal is important in order to accurately determine the

OUR and consequently the biological heating rate H_b, it is not important for the accuracy of the other terms in the heat balance, where the dry influent and effluent molar mass flow rates can be assumed equal (Messenger *et al*, 1992).

2.4.7 The Oxygen Supply Rate (Air + Pure Oxygen)

The mass flow rate of pure oxygen $M(O_2)_{in}^{O2}$, injected into the sludge recirculation line to supplement the aeration system during phase II, was monitored by a rotameter flow meter. Hourly readings were taken by the operating staff and recorded on the daily logsheet. The oxygen supply rate attributable to the pure oxygen OSR^{O2} is therefore given by:

$$OSR^{O2} = \frac{M(O_2)_{in}^{O2}}{184} = 0.00543\,M(O_2)_{in}^{O2} \quad \text{...kg}(O_2)/\text{m}^3.\text{h} \quad (2.5)$$

The overall oxygen supply rate OSR^{AIR+O2} during oxygen supplementation is obtained by combining Eq's 2.4 and 2.5 viz:

$$OSR^{AIR+O2} = OSR^{AIR} + OSR^{O2} = 0.00151\,Q(AIR)_{in} + 0.00543\,M(O_2)_{in}^{O2}$$

$$\text{...kg}(O_2)/\text{m}^3.\text{h} \quad (2.6)$$

Rotameter oxygen supply rates were regularly cross checked with oxygen content readings on the bulk oxygen storage vessel (6000kg(O_2) capacity).

2.4.8 Aerobic Reactor Gas Stream Humidity

No measurements of the humidity of the influent air stream or effluent gas stream were made. Instead, to allow calculation of the water vapour heat loss rate H_v, it was assumed that; (1) the influent air stream was saturated with water vapour after passage through the liquid ring compressor - significant quantities of water were continually being drained from the influent gas stream pipeline; and (2) the effluent gas stream was saturated with water vapour after passage through the sludge in the reactor - significant quantities of condensate were observed on the walls and roof of the reactor head space. The close correlation between the specific heat yield Y_h (defined as the quantity of heat generated biologically per unit mass of oxygen utilised MJ/kg(O_2)) obtained in this study with that determined by Messenger *et al* (1992), who did measure vent gas water vapour flow rates, indicated that these assumptions were reasonable.

2.4.9 The Respiration Quotient

The respiration quotient Y_{CO2} is defined as the number of moles of carbon dioxide generated per mole of oxygen utilised. The expression for Y_{CO2}, when oxygenation is with air only (derived in Section 3.2 below from the aerobic reactor gas stream mass

balance), in terms of the carbon dioxide $\%(CO_2)_{out}$ and oxygen $\%(O_2)_{out}$ dry effluent gas volumetric fractions is as follows:

$$Y_{CO_2} = \frac{\%(CO_2)_{out}\left(100 - \%(O_2)_{in}\right)}{\%(O_2)_{in}\left(100 - \%(O_2)_{out} - \%(CO_2)_{out}\right) - \%(O_2)_{out}\left(100 - \%(O_2)_{in}\right)}$$

$$\ldots \text{mol}(CO_2)_{gen}/\text{mol}(O_2)_{ut} \quad (3.27)$$

During Phase I, the carbon dioxide and oxygen fractions in the aerobic reactor effluent gas stream were monitored approximately three times weekly during each steady state period using a combustion analyser (which also gave the effluent gas temperature). By measuring both the oxygen and carbon dioxide fractions, an accurate estimation could be made of the rate of oxygen utilisation ***OUR*** ($kg(O_2)/m^3.h$), without having to measure the total vent gas volumetric flow rate (see Section 3.2.7 below).

It should be noted that sometimes in heat balance calculations on thermophilic aerobic reactors, in particular those using air, the effluent molar gas flow rate is assumed equal to the influent molar gas flow rate, which assumes that the respiration quotient Y_{CO2} is unity. This assumption obviates having to measure the effluent gas molar (or volumetric) flow rate, and the oxygen utilisation rate is calculated from only two measurements i.e. the influent gas flow rate and the effluent gas (% v/v) oxygen concentration. However if Y_{CO2} is not unity, as has been found by Messenger *et al* (1992) then this approach leads to considerable error (up to 20%) in the calculated ***OUR***. With an additional measurement i.e. the carbon dioxide volumetric fraction in the effluent gas, it is neither necessary to assume unity for the Y_{CO2} nor measure the effluent gas volumetric flow rate to obtain an accurate estimation of the oxygen utilisation rate ***OUR*** (see Section 3.2.7 for the derivation of the ***OUR*** equation).

During Phase II, when oxygenation was with both air and pure oxygen, no suitable instrument was available to measure the vent gas carbon dioxide concentration. Consequently, it was decided to accept the value for Y_{CO2} of 0.70 $\text{mol}(CO_2)/\text{mol}(O_2)$ determined during phase I, to allow the gas component mass balances to be resolved (see Section 3.3 below). Acceptance of the 0.70 value for phase II was considered valid due to (1) the high degree of precision obtained in the measurement of this value during phase I (see Section 4.3.6) and (2) the close agreement with the value of 0.67 determined by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor.

2.4.10 The Oxygen Utilisation Rate (Air Only)

The rate of oxygen utilisation ***OUR*** is defined as the mass of oxygen utilised by the bacteria per unit process volume per hour ($kg(O_2)/m^3.h$). Because of the fundamental

importance of the rate of oxygen utilisation ***OUR*** to the performance of the aerobic reactor, this aspect is dealt with in detail in Chapter 3.[4] The expression for ***OUR***,when oxygenation is with air only (derived in Section 3.2), in terms of (1) the volumetric air flow rate into the reactor $Q(AIR)_{in}$, (2) the oxygen fraction $\%(O_2)_{out}$ in the effluent gas stream, and (3) the respiration quotient Y_{CO2} (see Section 2.4.8 above) is as follows:

$$OUR = \frac{m_w(O_2)\ \rho(AIR)\ Q(AIR)_{in}\left(\%(O_2)_{in} - \%(O_2)_{out}\right)}{m_w(AIR)_{in} V_p\left(100 - \%(O_2)_{out} + Y_{CO_2}\%(O_2)_{out}\right)} \quad ...kg(O_2)/m^3.h \quad (3.42)$$

The average Y_{CO2} value observed from 118 pairs of $\%(O_2)_{out}$ and $\%(CO_2)_{out}$ data during phase I was 0.70 $mol(O_2)_{gen}/mol(O_2)_{ut}$ (see Section 4.3.6). As discussed, this value compared favourably with the 0.67 value observed by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor. Because the spread in Y_{CO2} values observed was minimal (see Fig 4.6) a constant value of Y_{CO2} = 0.70 $mol(CO_2)_{gen}/mol(O_2)_{ut}$ could be accepted with confidence. Equation 3.34 therefore simplifies to:.

$$OUR = \frac{Q(AIR)_{in}}{138} \times \frac{\left(21 - \%(O_2)_{out}\right)}{\left(100 - 0.3 \times \%(O_2)_{out}\right)} \quad ...kg(O_2)/m^3.h \quad (3.44)$$

It should be noted that the use of Eq 3.44 above does not imply that the carbon dioxide fraction $\%(O_2)_{out}$ does not need to be measured. It does need to be measured to obtain an accurate measure of the Y_{CO2} value.

2.4.11 The Oxygen Utilisation Rate (Air + Pure Oxygen)

In order to be able to determine the oxygen utilisation rate (when oxygenation is with both air and pure oxygen) in terms of an overall rate OUR^{AIR+O2} and separate contributions from the oxygen in the air stream OUR^{AIR} and the oxygen from the pure oxygen stream OUR^{O2} , it is assumed that the diffused air aeration system and the pure oxygen injection system operate independently of each other, i.e. neither one influences the performance of the other (see Section 3.3.2). This was verified during phase II (see Sections 7.3.4 and 7.3.5). As such, each system is considered to contribute individually to the overall oxygen utilisation rate viz:

$$OUR^{AIR+O2} = OUR^{AIR} + OUR^{O2} \quad ...kg(O_2)/m^3.h \quad (3.50)$$

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[4] In chapter 3 the reader is introduced to the theory of autothermal thermophilic aerobic treatment and formulae for the rate of oxygen utilisation ***OUR***, the biological heating rate H_b and the other terms in the steady state heat balance are derived.

To enable the individual components OUR^{AIR} and OUR^{O2} of the overall oxygen utilisation rate OUR^{AIR+O2} to be determined, the oxygen concentration in the vent gas is first measured with pure oxygen injection taking place $\%(O_2)_{out}^{AIR+O2}$. The pure oxygen supply is then switched off. After waiting 15 minutes (sufficient time to allow the vent air stream to flush any O_2 and CO_2 derived from the pure oxygen stream, from the reactor head space) the oxygen concentration is measured again $\%(O_2)_{out}^{AIR}$. The expression for the oxygen utilisation rate attributable to the oxygen from the pure oxygen stream OUR^{O2} (derived in Section 3.3 below) is as follows:

$$OUR^{O2} = \frac{M(O_2)_{in}^{O2}}{184} - \frac{\%(O_2)_{out}^{AIR+O2}\left(1.25Q(AIR)_{in} + 0.7M(O_2)_{in}^{O2}\right)}{184\left(100 - 0.3\%(O_2)_{out}^{AIR+O2}\right)} + \frac{1.25\%(O_2)_{out}^{AIR}.Q(AIR)_{in}}{184\left(100 - 0.3\%(O_2)_{out}^{AIR}\right)}$$

...$kg(O_2)/m^3.h$ (3.88)

The expression for the oxygen utilisation rate attributable to the oxygen from the air stream OUR^{AIR} is as given by Eq 3.44 viz:

$$OUR^{AIR} = \frac{Q(AIR)_{in}}{138} \times \frac{\left(21 - \%(O_2)_{out}^{AIR}\right)}{\left(100 - 0.3 \times \%(O_2)_{out}^{AIR}\right)}$$

...$kg(O_2)/m^3.h$ (3.44)

The expression for the overall oxygen utilisation rate during oxygen supplementation OUR^{AIR+O2} (derived in Section 3.3 below) is as follows:

$$OUR^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{184} + \frac{0.279Q(AIR)_{in}}{184} - \frac{\%(O_2)_{out}^{AIR+O2}\left(1.25Q(AIR)_{in} + 0.7M(O_2)_{in}^{O2}\right)}{184\left(100 - 0.3\%(O_2)_{out}^{AIR+O2}\right)}$$

...$kg(O_2)/m^3.h$ (3.91)

2.4.12 The Oxygen Transfer Efficiency (Air Stream)

The oxygen transfer efficiency *OTE* is defined as the ratio of the oxygen utilisation rate and the oxygen supply rate expressed as a percentage i.e *OTE* = 100 x *OUR/OSR*. By combining equations Eq 3.48 (also Eq 2.4) for *OSR* and Eq 3.44 for *OUR*, the expression for the oxygen transfer efficiency attributable to the oxygen from the air stream, in terms of the oxygen fraction in the effluent gas stream $\%(O_2)_{out}$, is as follows:

$$OTE = \frac{\left(100 - 4.76\%(O_2)_{out}\right)}{\left(100 - 0.30\ \%(O_2)_{out}\right)} \times 100$$

...% (3.36)

2.4.13 The Oxygen Transfer Efficiency (Pure Oxygen Stream)

The oxygen transfer efficiency attributable to the oxygen from the pure oxygen stream OTE^{O2}, in terms of the oxygen fractions in the effluent gas stream with pure oxygen injection both on and off ($\%(O_2)_{out}^{AIR+O2}$ and $\%(O_2)_{out}^{AIR}$) and the two measurable parameters for the influent gas flow rates ($Q(AIR)_{in}$ and $M(O_2)_{in}^{O2}$), is obtained by combining equations Eq 3.92 (also Eq 2.5) for OSR^{O2} and Eq 3.88 for OUR^{O2} as follows:

$$OTE^{O2} = 100 - \frac{\%(O_2)_{out}^{AIR\cdot O2}\left(125Q(AIR)_{in} + 70M(O_2)_{in}^{O2}\right)}{M(O_2)_{in}^{O2}.\left(100 - 0.3\%(O_2)_{out}^{AIR\cdot O2}\right)} + \frac{125\%(O_2)_{out}^{AIR}.Q(AIR)_{in}}{M(O_2)_{in}^{O2}.\left(100 - 0.3\%(O_2)_{out}^{AIR}\right)}$$

...% (3.95)

2.4.14 The Aerobic Reactor Foam Level

The foam level on the surface of the sludge in the aerobic reactor was measured at hourly intervals by the operating staff using a dipstick inserted through an inspection hole in the roof of the reactor.

2.4.15 The Aerobic Reactor Heat Balance Terms

The heating performance of the aerobic reactor was assessed by establishing around the reactor a heat balance in which the heat sources and heat sinks at steady state conditions are equated (see Section 3.1.8). The overall steady state heat balance can be expressed bas follows:

$$H_b + H_m = H_s + H_v + H_g + H_w \quad \text{...MJ/h} \quad (3.8)$$

where:

H_b = Rate of biological heat generation
H_m = Rate of heat energy input from the mixing device
H_s = Rate of sensible heat loss with the sludge leaving the reactor
H_v = Rate of vapour heat loss with the effluent gas stream
H_g = Rate of sensible heat loss with the effluent gas stream
H_w = Rate of heat loss through the walls of the reactor

During phase I, all the terms in the steady state heat balance except H_b were determined from various measurements taken in and around the aerobic reactor (see Table 2.5) and an application of the derived formulae listed below. The theory and development of these formulae is contained in Chapter 3.

The Water Vapour Heat Loss Rate in the Effluent Gas (Eq 3.129)
(Oxygenation with Air Alone)

$$H_v^{AIR} = 2.38 \times 1.205\, Q(AIR)_{in} \left(\frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{0.62 \times alog_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)}{1240 - alog_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)} \right)$$

The Effluent Gas Sensible Heat Loss Rate (Eq 3.151)
(Oxygenation with Air Alone)

$$H_g^{AIR} = 1.21\, Q(AIR)_{in} \left(0.001\left(T(AIR)_{out} - T(AIR)_{in}\right) + \frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} 0.00187\, T(AIR)_{out} \right)$$

The Sludge Sensible Heat Loss Rate (Eq 3.161)
(Oxygenation with Air Alone)

$$H_s^{AIR} = 4.04 \left(\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{1.21\, T(SL)_r Q(AIR)_{in}}{1000} \times \frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{Q(SL)_{in} T(SL)_{in}}{24} \right)$$

The Rate of Heat Loss from the Walls of the Reactor (Eq 3.195)
(Valid for Oxygenation with Air Alone and with Air + Pure Oxygen)

$$H_w = 0.82\left(T(SL)_r - T(SL)_d\right) + 0.49\left(T(SL)_r - T(AIR)_{in}\right)$$

The Rate of Mechanical Heat Input to the Reactor (Eq 3.200)
(Valid for Oxygenation with Air Alone and with Air + Pure Oxygen)

$$H_m = 1.65 \times I$$

where all the terms H_v^{AIR}, H_g^{AIR}, H_s^{AIR}, H_w and H_m have units MJ/h. The symbols are defined in Table 2.5 below.

Once the values for each of the heat balance terms have been determined, the biological heating rate H_b can be calculated by difference. A computer programme, written to perform the necessary computation, is listed in Appendix 8.

Table 2.6 Monitoring Parameters Required for Solving the Steady State Heat Balance

Symbol	Description of Monitoring Parameter	
$Q(AIR)_{in}$	Dry air volumetric flow rate into the reactor	m^3(STP)/h
$T(AIR)_{in}$	Influent gas stream temperature to the reactor	°C
$T(AIR)_{out}$	Effluent gas stream temperature from the reactor	°C
$T(SL)_r$	Temperature of the sludge leaving the reactor	°C
$T(SL)_{in}$	Temperature of the feed sludge to the reactor	°C
$Q(SL)_{in}$	Influent feed sludge flow rate to the reactor	m^3/d
$T(SL)_d$	Temperature of the sludge in the digester	°C
I	Current drawn by the mixing pump	amperes

The biological heating rate H_b is linked to the oxygen utilisation rate ***OUR*** via the specific heat yield Y_h (see Section 3.9) viz.

$$H_b = Y_h \, OUR \, V_p \qquad \text{(where } V_p \text{ = Process Volume)} \qquad \text{...MJ/h} \quad (3.4)$$

Once the biological heating rate H_b is known, by applying Eq. 3.4, the specific heat yield Y_h can be determined for the experimental system. For design, knowing Y_h allows H_b to be determined directly by measuring the oxygen utilisation rate ***OUR***. With H_b known via the ***OUR***, each term in the heat balance can be calculated directly, and the heat balance equation (Eq. 3.8) can be used for design purposes (see Chapter 5), where usually the unknown to be determined is the retention time (or sludge loading rate) to achieve a desired reactor temperature.

For phase II, with pure oxygen supplementation taking place, the heat balance terms for the water vapour heat loss ($H_v \rightarrow H_v^{AIR+O2}$), the gas sensible heat loss ($H_g \rightarrow H_g^{AIR+O2}$), and the sludge sensible heat loss ($H_s \rightarrow H_s^{AIR+O2}$), were modified to take into account the increased gas flow rate through the system. The revised equations (derived in Chapter 3) are as follows:

The Water Vapour Heat Loss Rate in the Effluent Gas (Eq 3.133)
(Oygenation with Air + Pure Oxygen)

$$H_v^{AIR+O2} = 2.38\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left(\frac{0.62\,a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{0.62\,a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}{1240-a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}\right)$$

The Effluent Gas Sensible Heat Loss Rate (Eq 3.153)
(Oygenation with Air + Pure Oxygen)

$$H_g^{AIR+O2} = \left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left[0.001\left(T(AIR)_{out} - T(AIR)_{in}\right) + \frac{0.62\,a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}\,0.00187\,T(AIR)_{out}\right]$$

The Sludge Sensible Heat Loss Rate (Eq 3.163)
(Oygenation with Air + Pure Oxygen)

$$H_s^{AIR+O2} = 4.04\left[\frac{T(SL)_r\,Q(SL)_{in}}{24} - \frac{T(SL)_r\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right)}{1000}.\frac{0.62\,a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{Q(SL)_{in}\,T(SL)_{in}}{24}\right]$$

The measurable parameter required to resolve the above heat balance terms when oxygenation is with both air and pure oxygen (additional to those listed in Table 2.5 above) is the mass flow rate of pure oxygen $M(O_2)_{in}^{O2}$ injected into the sludge recirculation line (see Section 2.4.7 above for details).

2.4.16 The Breakdown of Organic Matter

To determine the breakdown of biodegradable organic matter across each stage in the dual digestion process, the concentration of total solids (TS), volatile solids (VS) and the chemical oxygen demand (COD) were determined on the feed sludge, aerobic sludge and anaerobic sludge on a daily basis. These tests were carried out on the samples as they were taken, without prior dewatering. All three methods are as described in Standard Methods (1989). The reduction in the aerobic reactor outflow water volume due to water vaporisation (about 5 to 7%) was not taken into account in calculating the % removals.

Method for Determining the Total Solids and Volatile Solids Concentration

A measured volume of sample is dried in a weighed crucible at 105°C. The crucible plus sample is then cooled in a desiccator and weighed to determine the total solids content. The crucible plus dried sample is then incinerated at 600°C for one hour and following

cooling in a desiccator again, is reweighed to determine the inorganic ash (IS) content. Both the total solids TS and volatile solids VS (TS minus IS) are reported in units of g/l (equivalent to kg/m^3)

Method for Determining the Chemical Oxygen Demand

The sample is macerated, diluted 100 times, and a volume of 10ml taken for the COD test. The diluted sample is refluxed in concentrated sulphuric acid with a known excess of potassium dichromate. After digestion, the remaining unreduced dichromate is titrated with ferrous ammonium sulphate to determine the amount of dichromate consumed. The concentration of oxidisable organic matter in the sample is calculated in terms of the oxygen equivalent and reported as g(O)/l.

2.4.17 The Conditioning effects of Aerobic Pre-treatment

To determine the conditioning effects of aerobic pre-treatment the pH, ammonium ion concentration ((NH_4^+)), conductivity (COND), volatile acid alkalinity (VAA), and bicarbonate alkalinity (BA) were determined on the feed, aerobic, and anaerobic sludges on a daily basis.

Determination of pH

The pH of the sample is measured with a pH meter which is calibrated with standard buffer solutions of pH 7 and pH 4.

Determination of Ammonium Ion Concentration

The samples, after the appropriate dilutions, is buffered at pH 9.5 with borate buffer and distilled into a solution of boric acid. This acid solution is then titrated with standard sulphuric acid solution using mixed indicator. The ammonium ion concentration is reported as g(N)/l. The feed sludge samples were diluted ten times, aerobic sludge samples twenty times, and the anaerobic sludge samples fifty times. The method is as described in Standard Methods (1989).

Determination of Bicarbonate and Volatile Acid Alkalinity

The sample is centrifuged at 6000rpm for 15 minutes. A volume of 50ml of sludge supernatant is then poured into a glass evaporating dish. The supernatant sample is titrated with N/2 sulphuric acid solution to a pH of 4.0 using a pH meter. Further sulphuric acid is added to bring the pH to 3.5. The sample is then simmered gently on a hot plate for 3 minutes and allowed to cool to room temperature. The pH of the sample is then brought back to pH 4.0 by adding N/20 sodium hydroxide solution. The sample is the titrated back to pH 7.0 with the sodium hydroxide solution. The total alkalinity is calculated from the sulphuric acid titration and the volatile acid alkalinity

is calculated from the sodium hydroxide titration. The bicarbonate alkalinity is obtained from the difference between the two results. All alkalinities are reported as mg($CaCO_3$)/l. This method for determining the Alkalinities is described by Dilallo and Albertson (1962).

2.4.18 Sludge Disinfection

To monitor the disinfecting capabilities of aerobic pre-treatment, faecal coliform and ascaris ova determinations were performed on grab samples of the three sludge types i.e. feed, aerobic reactor, and anaerobic digester, taken at regular intervals.

Determination of Faecal Coliform Concentration

The sample is diluted using sterile buffered distilled water. Dilution can range from 10^1 to 10^8 depending on the type of sample. The diluted sample is then filtered through a 0.45µm membrane filter, which is then placed in a petri dish containing M-FC agar. The dish is then incubated at 44.5°C for 24hrs, after which the dish is examined under low magnification. All blue colonies which have developed are counted. The faecal coliform concentration is expressed as faecal coliforms per 100ml(or cl). The method is as described in Standard Methods (1989).

Determination of Viable Ascaris Ova Concentration

A specific volume of sludge, dependent upon the solids concentration, is passed through a Vissor-filter to remove the bulk of the sludge solids. The filtrate is then passed through a 8µm membrane filter which retains the ascaris ova on the filter. After drying at room temperature, a thin layer of immersion oil is spread onto the filter and the total number of ascaris ova are counted under a microscope. The viability of the ascaris ova is determined by inoculating the filtrate from the Vissor-filter for 6 weeks at 20°C and then counting the viable and non-viable ova retained on the membrane filter under the microscope. The concentration of ascaris ova is expressed as the total number of ova per 1 gm of dried sludge and the viability as the percentage viable ova with respect to the total count. The method is as described by Morrison (1986)

2.4.19 The Dewaterability of the final Sludge

The dewaterability of the final sludge from the anaerobic digester was monitored by measuring the specific resistance to filtration (SRF) on grab samples at frequent intervals. As the test is only semi-qualitative, in that the results do not directly measure sludge dewaterability, comparison is made with SRF values obtained with other sludge types.

Determination of the Specific Resistance to Filtration

A wet piece of Whatman 541 filter paper is placed into a Buchner funnel under pressure. The funnel is then filled with 200ml of sample of known concentration and a controlled constant vacuum is applied. The filtrate volume is then recorded as a function of time. The specific resistance to filtration (SRF) is calculated from a derivation of D'Arcy's Law and is reported as m/kg. The method is as described by Swanwick *et al* (1961) and Smollen (1986).

2.4.20 The Stability of the Final Sludge (Phase II Only)

During phase II, the anaerobic digester operated at thermophilic temperatures (50°-55°C) and during the latter part of the period the hydraulic retention time was comparatively short (<10 days). It was therefore decided to test the stability of the final sludge, obtained during this period, by aerating a fixed volume (3ℓ) sampled from the digester and measuring the change in the oxygen utilisation rate with time.

Certain sludge guidelines (see Ekama, 1992) state that a sludge is required to have a specific ***OUR*** (***SOUR***) of less than 1mg(O)/g(TSS).h to be regarded as stable. Owing to the difficulty of interpreting this specification, i.e. over what time period must the ***SOUR*** be measured and at what stage in the test must the ***SOUR*** be <1mg(O)/g(TSS).h, a number of sludges from different sludge treatment systems were tested by measuring their ***SOUR*** over a four day (96hrs) period. These results are discussed in Section 7.4.4 below and presented in full in Appendix 9.

2.5 LABORATORY SCALE EVALUATION OF THE CONDITIONING EFFECTS OF AEROBIC TREATMENT

2.5.1 Motivation for the Study

At the long aerobic reactor retention times (3-6 days) required for air oxygenation, it was not possible to reduce the retention time in the anaerobic digester sufficiently to verify the claim made for the dual digestion process that aerobic autothermal thermophilic pre-treatment reduces the required anaerobic digestion time from 25d to 10d. Consequently it was decided to test the claim at laboratory scale.

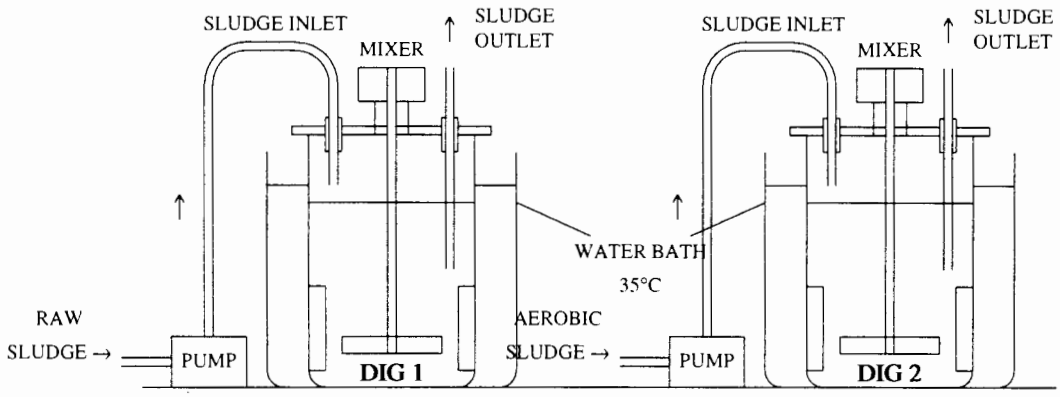

Figure 2.4 Schematic of the Laboratory Scale Digesters

2.5.2 Experimental Procedure

Two laboratory scale digesters (Digesters 1 and 2) (see Figure 2.4) were operated for a two month period (June-July 1990). Both digesters were operated at 8 day retention times. Digester 1 was fed raw primary sludge from the gravity thickener, whilst Digester 2 was fed aerobic reactor effluent sludge. The feed sludge for the two systems was obtained from the composite samples of the feed and aerobic reactor sludges of the full scale system. The results of the analysis of these samples for the full scale system were used also for the laboratory scale systems (for the tests conducted see Section 2.4.1). For convenience these feed sludge results are tabulated also with the laboratory digester results in Appendix 2.

Both digesters were kept in a water bath with the temperature controlled at 35°C. The hydraulic capacity of each digester was 6 litres. To obtain the desired retention time of 8 days, 250ml of feed sludge were added every 8 hours with a peristaltic pump to give a daily loading rate of 0.75 l/d. Both digesters were continually mixed with an electrical motorised mixer. The digesters were sealed to prevent air ingress. Desludging took place once daily i.e. 0.75 l was abstracted daily prior to the daytime feeding period. Samples were taken three times weekly from the abstracted sludge and tested for TS, VS, (NH_4^+), Bicarbonate and Volatile Acid Alkalinity. Gas composition and production were not measured, and digester stability or failure was assessed on the results of the measured sludge parameters. The results are listed in Appendix 3 and discussed in Section 4.5.

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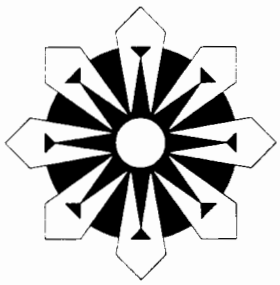

CHAPTER 3

THERMOPHILIC AEROBIC TREATMENT

3.1 THE BASICS OF THERMOPHILIC AEROBIC TREATMENT

3.1.1 Introduction

The biological aerobic treatment of sewage sludge, produces (1) a range of by-products (largely carbon dioxide and water) from the oxidative degradation of the organics in the sludge, (2) new biomass through bacterial synthesis, and (3) heat. It is through the exploitation of the biological heat produced that certain aerobic treatment systems such as the Autothermal Thermophilic Aerobic Digestion (ATAD) process and the autothermal aerobic reactor in the Dual Digestion process are able to operate in the thermophilic temperature range.

At thermophilic temperatures biological reactions proceed much faster than at ambient temperature. For the ATAD process this means that shorter retention times, and thus lower capital costs, are possible to achieve the same level of stabilisation and much higher pathogen reductions in comparison to the cold aerobic digestion process. In the aerobic reactor of dual digestion, the biological heat which is generated is used principally to raise the sludge temperature in order to effect disinfection and conditioning with as little sludge stabilisation as possible; stabilisation takes place principally in the anaerobic digester.

In order to be able to (1) provide the necessary information with regard to the design of the aerobic reactor, and (2) predict operating temperatures under a variety of different loading conditions for installed reactors, the biological heat generation rate in the process needs to be quantified. To do this consideration must be given to the various factors which influence it, namely:

- The biological reactions which take place
- The rate at which these biological reactions proceed
- The stoichiometry of biological heat generation
- Conditions which limit biological activity

- The conservation of heat (the steady state heat balance)
- The application of heat transfer theory into practice

These aspects are discussed in this Chapter with the view of developing equations to quantify the water mass, oxygen mass and heat balances for the aerobic reactor aerated with air.

3.1.2 Biological Reactions Occurring under Aerobic Conditions

The biological reactions which occur during the aerobic treatment of sewage sludges are extremely complex and varied and the intricacies of such reactions are poorly understood. However, by accepting a simplistic view, these biological reactions can be schematically represented as shown in Figure 3.1.

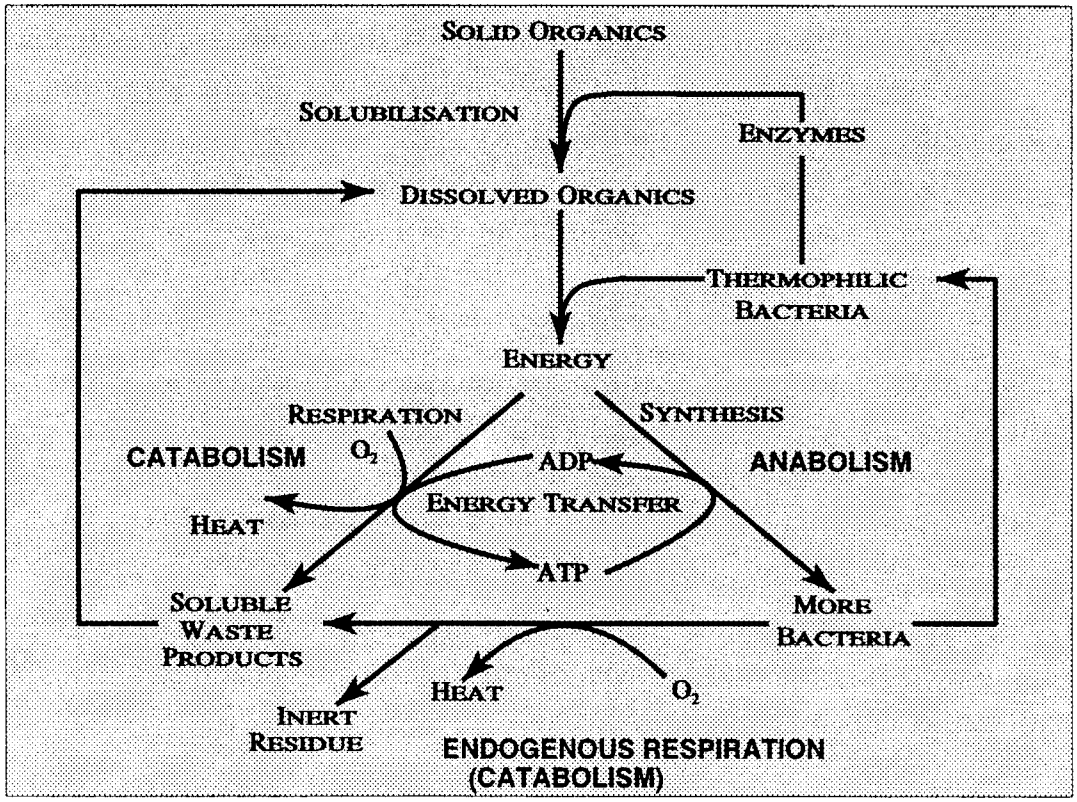

Figure 3.1 Biological Reactions Occurring under Aerobic Conditions

The organic solids in the feed sludge are first *solubilised* by the action of enzymes excreted by the thermophilic bacteria. The soluble organic material, consisting mainly of reduced organic acids, is then utilised as substrate for growth by the thermophilic bacteria. Degradation of the organics through respiration yields *energy* and a part of this energy is retained through *synthesis* of new bacterial mass, the remainder is lost as heat. Besides growth there is also a reduction in bacterial mass through *endogenous respiration*. Degradation of the live bacterial mass provides energy for cell maintenance. As the

bacterial mass is not completely biodegradable, a fraction of the original mass remains as inert endogenous residue. In endogenous respiration the bacteria also are not 100% efficient at utilising the energy gained through endogenous respiration, a portion of this energy is also released as *heat*. During the breakdown of the organic matter both carbon dioxide and ammonia are released and this influences the alkalinity of the liquid sludge; the effects of these two products on the alkalinity is discussed in Section 3.1.9. Generally as the temperature increases so the biological activity and kinetic rates increase both for the growth and endogenous processes.

In terms of the above conceptual behaviour in thermophilic aerobic processes, the rate at which the biological processes take place will depend on a number of factors such as:

- The concentration of biodegradable organic matter
- The temperature of the system
- The mixing efficiency and oxygen transfer rate
- The nature of the substrate
- The activity of the thermophilic bacteria

A function to describe the rate of biological heat generation and oxygen utilisation, which considers all the factors listed above would be extremely complex and difficult to apply in practice, so a simplified empirical approach is adopted in modelling the ATAD process.

3.1.3 Modelling the ATAD Process

In the ATAD process where the objective of sludge treatment is stabilisation it is necessary to quantify the rate of volatile solids destruction in order to calculate the required process retention time to achieve a specified VS removal (e.g. 40%). Because VS reduction is the principal objective, which generally requires fairly long retention times (>5d), modelling of the process is based on a VS degradation approach. At the long retention times, the overall rate of volatile solids degradation by the bacteria can be adequately described by first order kinetics. At steady state conditions, with no deficiency of oxygen, the rate of volatile solids degradation or destruction was modelled by Andrews and Kambhu (1971) with a first order rate equation with respect to the biodegradable VS concentration (Eq 3.1) and the rates of COD removal, oxygen utilisation and heat generation were all linearly linked to the rate of VS destruction via empirical constants. i.e.

$$M(VS)_{dest} = \frac{V_p}{R_h}([BVS]_{in} - [BVS]_{out}) = -k_d.[BVS]_{out}.V_p \quad \text{...kg(VS)/h} \quad (3.1)$$

where:

$M(VS)_{dest}$ = The rate of VS destruction (kg(VS)/h)
k_d = The VS destruction rate coefficient (/h)
R_h = The hydraulic retention time (h)
V_p = The effective process volume (m^3)
$[BVS]_{in}$ = The biodegradable VS concentration in the *influent* sludge (kg/m^3)
$[BVS]_{out}$ = The biodegradable VS concentration in the *effluent* sludge (kg/m^3)

As the kinetic rates of the bacteria increase with increasing temperature, the rate of volatile solids destruction and hence the rates of oxygen utilisation and heat generation, are higher at thermophilic temperatures. The increase in rate is reflected in the above equation through a change in the rate coefficient k_d. The dependency of the rate coefficient k_d with temperature can be described using an Arrhenius equation:

$$k_T = k_{20}.\theta^{T-20} \quad .../h \quad (3.2)$$

Typically the rate of destruction doubles between 25°C and 50°C, illustrating the benefit of performing aerobic digestion in the thermophilic temperature range. Once the temperature increases above 60°C the decay rate decreases as the thermophilic organisms become inhibited by temperature.

For the aerobic reactor in dual digestion, the objective is not stabilisation but disinfection and conditioning. The design of the aerobic reactor is therefore not based on achieving a specific VS reduction but rather on achieving disinfection temperatures. In addition, Messenger *et al* (1992) found it inappropriate to employ the above kinetics (based on VS destruction) for modelling the aerobic reactor in dual digestion because:

- VS reduction is not an objective, and at the short retention times of the aerobic reactor compared to ATAD, large quantities of heat can be generated without significant VS reduction.

- VS destruction is not the dominant bacterial activity. Because of the shorter retention times (<2d) the feed sludge provides a significant quantity of external substrate for the bacteria. As a result, growth (synthesis) of new bacterial mass will constitute a large part of the bacterial activity which takes place within the reactor.

- The relatively high concentration of external substrate in the aerobic reactor sludge means that the biological oxygen demand is higher. Consequently the reactor is not necessarily operated under conditions of oxygen excess. Under oxygen limiting conditions, the rate of volatile solids destruction will be controlled by the rate of oxygen transfer into solution.

- The volatile solids test is prone to significant variability when dealing with sewage sludges. The actual percentage reduction in volatile solids across the reactor is relatively small (<5%) and to a large degree masked by the errors inherent in the volatile solids test.

For the air aerobic reactor in the Athlone dual digester, the objective is, as for the pure oxygen reactor, to achieve disinfection and conditioning. However, because longer retention times are required (3-5d) when using air, due to the cooling effects of passing large volumes of nitrogen through the process, a significant degree of volatile solids destruction does take place (approximately 25%) in the aerobic reactor. Whilst VS destruction is not a design parameter, in that it is not employed to quantify the biological heat generation, the VS kinetics are useful to predict the minimum VS concentration to ensure that there is no limitation on available substrate (see Section 4.4.2).

3.1.4 Biological Heat Generation

Biological heat generation is the fundamental operating criterion for the aerobic reactor in the dual digestion system. To be able to predict the potential for a particular aerobic reactor system to generate sufficient biological heat to sustain operation at disinfection temperatures in the thermophilic temperature range under a variety of differing operating conditions, it is important to be able to quantify the rate of biological heat generation.

Messenger *et al* (1992) found that the biological heat generation rate was directly proportional to the rate of oxygen utilisation by the bacteria. The constant of proportionality is the specific heat yield (Y_h) and was measured to be approximately 13.0 MJ/kg(O_2), a value which conforms closely to that obtained from bioenergetics and thermodynamics.

$$H_b = Y_h . M(O_2)_{ut} = 13.0 \times M(O_2)_{ut} \quad \text{...MJ/h} \quad (3.3)$$

where:

H_b = Biological heat generation rate (MJ/h)

Y_h = Specific heat yield coefficient in terms of oxygen utilisation (MJ/(kg(O_2))

$M(O_2)_{ut}$ = Mass rate of biological oxygen utilisation (kg(O_2)/h)

For the ATAD process, biological heat generation kinetics have generally been presented as a function only of volatile solids removal kinetics. The quantity of heat generated has been measured to be approximately 21 MJ/kg of volatile solids destroyed (Andrews and Kambhu, 1971).

3.1.5 The Supply, Transfer and Utilisation of Oxygen in the Aerobic Reactor

Accepting that the rate of utilisation of oxygen by the bacteria is the most appropriate parameter for determining the rate of biological heat generation, not only in the pure oxygen aerobic reactor but also in the air aerobic reactor, it is necessary to define a number of terms to describe the *supply, transfer and utilisation* of oxygen in the aerobic reactor.

With regard to the aeration system:

- The overall mass flow rate of oxygen to the aerobic reactor, defined as the oxygen supply rate, is given by $M(O_2)_{in}$ (kg(O_2)/h) and as per unit reactor volume by ***OSR*** (kg(O_2)/m^3.h).

- The mass flow rate of oxygen transferred into the sludge liquid per unit reactor volume, defined as the oxygen transfer rate, is given by ***OTR*** (kg(O_2)/m^3.h).

- The mass of oxygen transferred into the sludge liquid as a fraction of the mass of oxygen supplied to the aerobic reactor ***(OTR/OSR)*** is defined as the oxygen transfer efficiency ***OTE***.

With regard to the biological activity within the sludge:

- The mass rate at which the bacteria utilise the available oxygen, defined as the oxygen utilisation rate, is given by $M(O_2)_{in}$ (kg(O_2)/h) and as per unit reactor volume by ***OUR*** (kg(O_2)/m^3.h).

3.1.6 Limiting Conditions for Biological Oxygen Utilisation

In the aerobic reactor, the amount of heat generated by the thermophilic aerobic bacteria is *limited* by either the oxygen transfer rate ***OTR*** of the aeration system or the maximum oxygen utilisation rate of the bacteria OUR_{max}. To produce sufficient biological heat for the aerobic reactor to reach thermophilic temperatures an aeration system which can generate an appropriately high ***OTR*** is required. In addition, an adequate concentration of substrate and bacteria are necessary to ensure that the biological oxygen utilisation rate OUR_{max} does not inhibit the system from reaching thermophilic temperatures.

Providing sufficient substrate and thermophilic bacteria are available such that the reactor is not biological activity or OUR_{max} limited, the amount of oxygen available to the bacteria will be limited by the oxygen transfer rate ***OTR*** of the aeration system. Under such conditions the bacteria utilise all the available oxygen directly as it is transferred into solution and oxygen limiting conditions prevail within the sludge, i.e. the biological oxygen utilisation rate ***OUR*** is constrained to a value lower than the maximum OUR_{max} by the oxygen transfer rate ***OTR*** and ***OUR = OTR***. The dissolved oxygen concentration in the sludge is below detectable limits and the biological heat generation rate is given by:

$$H_b = Y_h.OUR.V_p = Y_h.OTR.V_p \quad \text{...MJ/h} \quad (3.4)$$

If there is a deficiency in either available substrate or bacterial population then the biological oxygen utilisation rate OUR_{max} may fall below the oxygen transfer rate ***OTR*** of the aeration system. Under such conditions oxygen utilisation is limited by the biological oxygen utilisation rate OUR_{max} and oxygen surplus conditions prevail. The dissolved oxygen concentration in the sludge increases above detectable limits and the biological heat generation rate is at its maximum given by:

$$H_{b_{max}} = Y_h.OUR_{max}.V_p \quad \text{...MJ/h} \quad (3.5)$$

3.1.7 Optimum Conditions for Biological Oxygen Utilisation

Ideally, the aeration system should be capable of transferring sufficient oxygen into solution to match or exceed the maximum biological oxygen utilisation rate which can occur within the reactor, i.e.

$$OTR_{max} > OUR_{max} \quad \text{...kg(O)/m}^3\text{.h} \quad (3.6)$$

Under normal operating conditions at Athlone (i.e. at feed sludge concentrations greater than 30g(TS)/l) the maximum oxygen transfer rate which could be generated by the aeration system was generally less than the maximum oxygen utilisation rate of the sludge. This was established by D.O. measurement in the reactor and finding an absence

of dissolved oxygen on most occasions . Under these conditions the maximum biological heat generation rate was limited by the maximum oxygen transfer rate of the aeration system i.e.

$$H_{b_{max}} = Y_h . OTR_{max} . V_p \quad \text{...MJ/h} \quad (3.7)$$

An ambient temperature (20°C) unsteady state aeration test with tap water was conducted (see Appendix 4 for details) and a maximum standard oxygen transfer rate of 30.5 kg(O_2)/h was measured, which for the 184 m^3 aerobic reactor, gives an OTR_{max} 0.166 kg(O_2)/m^3.h. However, this can vary significantly depending on conditions in the reactor, such as temperature, foam layer and mixing (see Sections 4.3.3 and 4.3.5). The test was conducted to obtain an approximate maximum value for the aeration system under standard conditions.

3.1.8 The Conservation of Heat

While external heating of sludges to thermophilic temperatures is generally regarded as uneconomical, the heat liberated in the biological oxidation of the biodegradable matter is sufficient to raise the sludge temperature into the thermophilic range, provided that this heat is contained to limit the heat losses.

Aerobic reactors are therefore designed on the basis of conserving thermal energy. Consequently it is necessary that the total heat production within the two reactors of the dual digester is sufficient to maintain the required temperatures in spite of the heat losses which occur to the ambient surroundings. The total heat sources into and heat losses from the system can be equated by performing a *heat balance* around the system.

The term *Heat Balance* refers to a special form of energy balance which has come into general use to describe thermal processes where changes in kinetic energy, potential energy, and work done are considered negligible. The principle of the heat balance is the conservation of energy (the first law of thermodynamics). The simplest statement of the principle for use in calculations centring around thermal processes is that, at steady state, "heat input = heat output".

The steady state heat balance is generally applied to processes which are continuously fed and the process is judged to be at steady state when the heat balance parameters, such as reactor temperature, do not change over time; implying that the heat sources (input) equal the heat losses (output). In the case of the Athlone aerobic reactor, feeding took place on a batch-wise basis at either 2 hourly or 4 hourly intervals. However, because the batch feed volumes were small in comparison with the reactor process volume (<3%), the change in reactor temperature before and after feeding was negligible

(<1°C). In addition, all data for the parameters monitoring the process were recorded as daily averages. In this manner diurnal fluctuations in air, feed sludge and ambient temperatures were smoothed. It is therefore acceptable to consider the aerobic process to be quasi-continuous and the steady state heat balance can be applied[1]. For the aerobic reactor (Figure 3.2), in the dual digestion system using air, the overall steady state heat balance can be expressed by:

$$H_b + H_m = H_s + H_v + H_g + H_w \quad \text{...MJ/h} \quad (3.8)$$

where:

H_b = Rate of biological heat generation
H_m = Rate of heat energy input from the mixing device
H_s = Rate of sensible heat loss with the sludge leaving the reactor
H_v = Rate of vapour heat loss with the effluent gas stream
H_g = Rate of sensible heat loss with the effluent gas stream
H_w = Rate of heat loss through the walls of the reactor

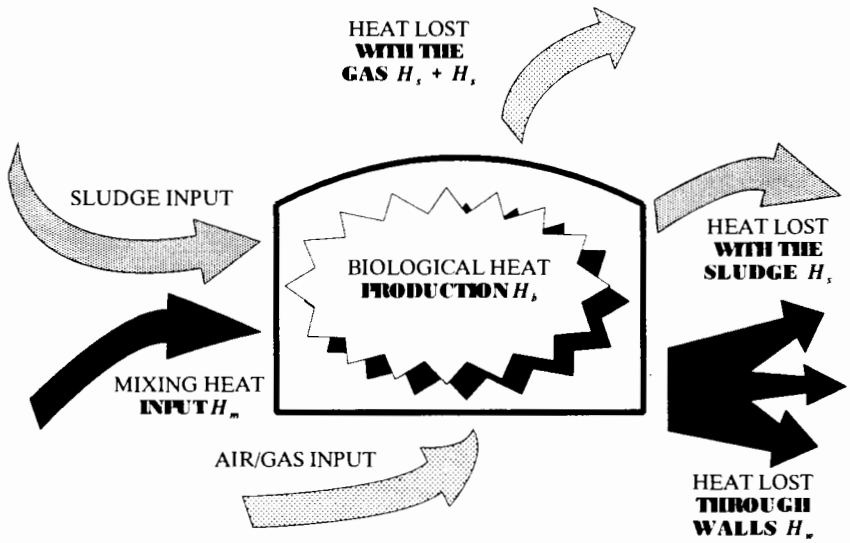

Figure 3.2 Heat Flows Around the Aerobic Reactor

All the above heat terms are expressed in units of MJ/h. Both sensible heat terms H_s and H_g are defined and calculated with respect to the influent sludge and air temperatures respectively. It is for this reason that the heat sources from the influent sludge and air are not included in Eq 3.8.

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[1] Unlike at Athlone, at Milnerton the aerobic reactor was batch fed on a draw and fill basis replacing 8.3% of the reactor volume per batch feed. This caused a 3°C drop in reactor temperature during filling. Consequently, an unsteady state heat balance had to be applied to the Milnerton aerobic reactor to calculate the different heat balance terms (Messenger *et al*, 1992).

3.1.9 Conditioning Effects of Aerobic Pre-Treatment

The conditioning effects brought on by aerobic pre-treatment in the reactor are primarily due to:

- The formation of organic acids
- The production of bicarbonate
- The production of ammonia alkalinity

In the anaerobic digestion system, it is claimed (Eastman and Ferguson, 1981) that the rate limiting step is the hydrolysis (solublisation) of particulate organic solids to organic acids. If this is so then the solublisation of these solids claimed to occur in the aerobic pre-treatment reactor in the dual digestion system should have the effect of reducing the minimum required retention time for the anaerobic digestion stage. It is interesting to note that Mason (1986) suggests that oxygen limitation in the aerobic reactor enhances the solublisation performance by increasing the production of hydrolysis enzymes.

The stability of the anaerobic digestion process is improved if there is an increase in the alkalinity of the feed sludge. The breakdown of organic matter in the aerobic pre-treatment reactor produces both carbon dioxide and ammonia. A large percentage of the carbon dioxide released during respiration by the bacteria, will enter the gas phase and pass out of the system. However, a fraction (<5%) remains in solution as bicarbonate as a result of the uptake of $\mathbf{H}^+$ ions in the ammonification of NH_3 (see Eq 3.10):

$$CO_{2(g)} + H_2O_{(aq)} \rightarrow H_2CO^*_{3(aq)} \rightarrow H^+_{(aq)} + HCO^-_{3(aq)} \quad \ldots (3.9)$$

The respiration quotient Y_{CO2} is defined as the number of moles of carbon dioxide generated per mole of oxygen utilised. Any carbon dioxide which passes into solution will have the effect of reducing the respiration quotient, but because this is relatively small the respiration quotient can be closely estimated from the CO_2 mass flow rate in the vent gas. The ammonia which is released via the breakdown of organics which contain N, pass into solution and take up protons to form ammonium ions. i.e.

$$NH_{3(aq)} + H^+_{(aq)} \rightarrow NH^+_{4(aq)} \quad \ldots (3.10)$$

The ammonium formed generally remains as such because it has been noted that an insignificant amount of nitrification takes place at thermophilic temperatures due to the die off of nitrifying bacteria in this temperature range. Furthermore, for the Athlone dual digester, only primary sludge, which ordinarily does not contain nitrifiers, was used as feed sludge. With the result that no nitrification took place in the aerobic reactor. Viewed simplistically therefore, it is the ammonium formed (i.e. the uptake of $\mathbf{H}^+$ ions) that produces the increase in alkalinity; the $\mathbf{H}^+$ ions taken up are replaced by the disassociation of dissolved CO_2 ($H_2CO_3^*$) leaving bicarbonate as depicted in Eq (3.9).

3.1.10 Application of Theory into Practice

The basic theory of biological oxygen utilisation, heat generation and volatile solids destruction during autothermal thermophilic aerobic treatment has been outlined above. The remainder of this Chapter deals specifically with applying this theory to the air oxygenated aerobic reactor in dual digestion. Formulae are developed, in terms of the measurable operating parameters (see Table 2.4), to enable the fundamental characteristics of the system to be quantified or estimated. Primarily this includes:

- The oxygen utilisation rate ***OUR***, oxygen transfer efficiency ***OTE*** and respiration quotient Y_{CO2}.
- The individual heat terms in the steady state heat balance, the specific heat yield coefficient Y_h and the biological heating rate H_b.
- The rate of volatile solids destruction $M(VS)_{dest}$ and the fraction of VS removal $f(VS)_{rem}^{O2lim}$
- The retention time R_h^{max} and feed solids concentration TS_{in}^{min} when substrate limitation will take place.
- The increase in ammonia concentration as a result of the breakdown in volatile solids

The derived formulae are applied in Chapter 4 where the results obtained from the investigation are presented, evaluated and discussed. They are used also in Chapter 5 where the performance of the aerobic reactor is modelled under different sets of operating conditions.

3.2 THE OXYGEN UTILISATION RATE (AIR ONLY)

3.2.1 Introduction

Phase I of the research programme investigated the performance of the aerobic reactor using air alone for oxygenation. The derivation of expressions to accurately estimate the supply, transfer, and utilisation of oxygen within the aerobic reactor under such conditions is based on the solution of the mass balances of the different components (nitrogen, oxygen, and carbon dioxide) in the dry gas stream, with the assumption that the gas behaves ideally. The measurable parameters selected for the purpose of solving the gas component mass balances were:

$Q(AIR)_{in}$ = The volumetric air flow rate into the aerobic reactor (m^3(STP)/h)
$\%(O_2)_{out}$ = The oxygen concentration in the dry effluent vent gas (% $^v/_v$)
$\%(CO_2)_{out}$ = The carbon dioxide concentration in the dry effluent vent gas (% $^v/_v$)

3.2.2 The Composition of the Influent and Effluent Gas Streams

During phase I of the investigation, the aerobic reactor of the Athlone dual digester was aerated with atmospheric air. Accordingly the composition of the influent gas stream is taken to be that of atmospheric air (Table 3.1).

Table 3.1 The Volumetric Composition of Dry Atmospheric Air

Constituent		Content (percent) by volume
Nitrogen	N_2	78.08
Oxygen	O_2	20.95
Argon	Ar	0.93
Carbon Dioxide	CO_2	0.03
Other Rare Gases	He, Kr, Xe	<0.01

It is assumed that the dry influent gas stream comprises of oxygen and nitrogen only with the oxygen content at the quoted value for atmospheric air (21.0 % $^v/_v$). The carbon dioxide content of the influent gas is assumed to be negligible in proportion to that generated in the reactor, which is true (<1%). Hence all the atmospheric fractions except oxygen (i.e. nitrogen, argon, carbon dioxide and the rare gases) are combined in, what is termed, an equivalent nitrogen fraction (79.0 % $^v/_v$). The composition of the influent gas stream is therefore as follows:

$$\%(N_2)_{in} + \%(O_2)_{in} = 100\% \quad \text{...\% v/v} \quad (3.11)$$

As the gas stream passes through the sludge in the aerobic reactor, a percentage of the oxygen supplied to the reactor is transferred into solution. The bacteria then utilise this oxygen and release carbon dioxide through respiration. The carbon dioxide exits the reactor in the effluent gas stream. The effluent gas stream therefore consists of three components:

$$\%(N_2)_{out} + \%(O_2)_{out} + \%(CO_2)_{out} = 100\% \quad \text{...\% v/v} \quad (3.12)$$

As the composition of the gas streams are defined in terms of (% $^v/_v$), a parameter which directly indicates the mole fraction, the formulae for the dry influent and effluent gas streams are initially derived in terms of **molar mass** flow rates denoted by n (kmol/h):

For the influent gas stream:

$$n(N_2)_{in} = n(AIR)_{in} \cdot \frac{\%(N_2)_{in}}{100} = n(AIR)_{in} \cdot \frac{(100 - \%(O_2)_{in})}{100} \quad \text{...kmol/h} \quad (3.13)$$

$$n(O_2)_{in} = n(AIR)_{in} \frac{\%(O_2)_{in}}{100} \quad \text{...kmol/h} \quad (3.14)$$

For the effluent gas stream:

$$n(N_2)_{out} = n(AIR)_{out} \frac{\%(N_2)_{out}}{100} = n(AIR)_{out} \frac{100 - \%(O_2)_{out} - \%(CO_2)_{out}}{100} \quad \text{...kmol/h} \quad (3.15)$$

$$n(O_2)_{out} = n(AIR)_{out} \frac{\%(O_2)_{out}}{100} \quad \text{...kmol/h} \quad (3.16)$$

$$n(CO_2)_{out} = n(AIR)_{out} \frac{\%(CO_2)_{out}}{100} \quad \text{...kmol/h} \quad (3.17)$$

where:

$n(i)_{in,out}$ = Molar mass flow rate of component *i* in the influent/effluent gas (kmol/h)
$n(AIR)_{in,out}$ = Molar mass (dry) flow rate of the influent/effluent gas stream (kmol/h)
$\%(i)_{in,out}$ = Vol. concentration of component *i* in the influent/effluent gas (% $^{v}/_{v}$)

A Summary of the dry gas stream composition, and the mass balances on the different components is given in Figure 3.3 below.

3.2.3 The Equivalent Nitrogen Gas Mass Balance

It is assumed that the equivalent nitrogen molar mass flow rate remains unchanged through the reactor i.e. there is negligible dissolution of nitrogen and other non-oxygen air components into the sludge.

$$n(N_2)_{in} = n(N_2)_{out} = n(AIR)_{out} \cdot \frac{\%(N_2)_{out}}{100} = n(AIR)_{in} \cdot \frac{\%(N_2)_{in}}{100} \quad \text{...kmol/h} \quad (3.18)$$

In terms of the measurable parameters $\%(O_2)_{out}$ and $\%(CO_2)_{out}$ the equivalent nitrogen gas mass balance can be expressed as follows:

$$n(AIR)_{out} \cdot \frac{(100 - \%(O_2)_{out} - \%(CO_2)_{out})}{100} = n(AIR)_{in} \cdot \frac{(100 - \%(O_2)_{in})}{100} \quad \text{...kmol/h} \quad (3.19)$$

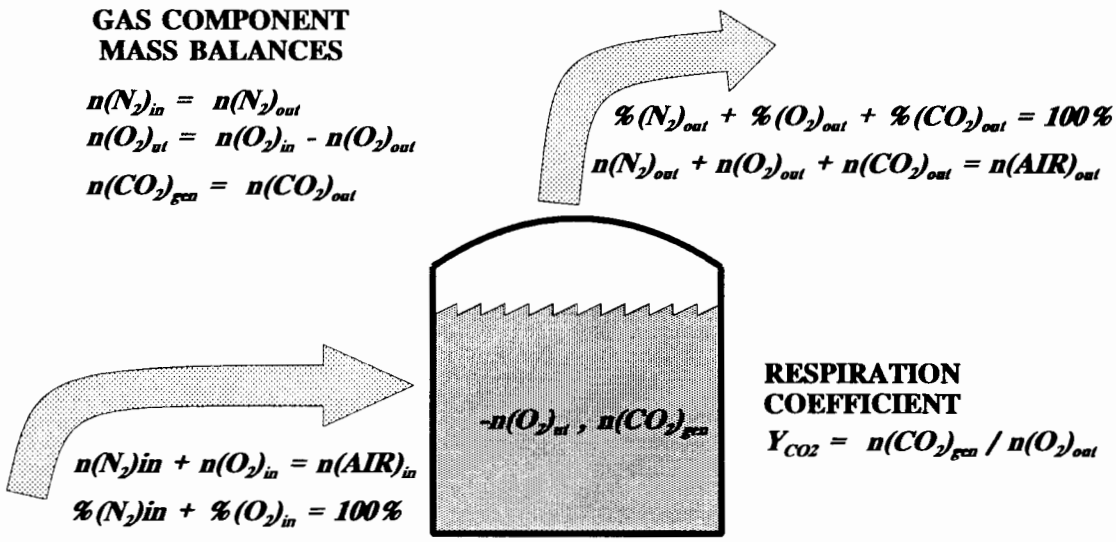

Figure 3.3 Gas Components Across the Aerobic Reactor (Oxygenation is with Air alone)

3.2.4 Relationship Between the Influent and Effluent Gas Streams

From the equivalent nitrogen gas mass balance, the relationship between the dry influent and effluent gas stream molar mass flow rates can be obtained by rearranging equation Eq 3.19 viz:

$$n(AIR)_{out} = n(AIR)_{in} \times \frac{(100 - \%(O_2)_{in})}{(100 - \%(O_2)_{out} - \%(CO_2)_{out})} \quad \text{...kmol/h} \quad (3.20)$$

By assuming that the molar mass flow rates of the dry influent and effluent air are not equal, enables the oxygen mass balance to be **accurately** calculated. Whilst this assumption is important for the accurate determination of the oxygen utilisation rate *OUR* and the biological heating rate H_b, it is not so important for accurately determining the sensible heat (H_s and H_g) and vapour heat (H_v) loss rates from the system, where it can be where it is assumed that the dry influent and effluent molar mass flow rates are equal.

3.2.5 The Oxygen Gas Mass Balance

The **influent** molar mass flow rate of oxygen to the aerobic reactor was given above by Eq 3.14 viz:

$$n(O_2)_{in} = n(AIR)_{in} \frac{\%(O_2)_{in}}{100} \quad \text{...kmol/h} \quad (3.14)$$

The **effluent** molar mass flow rate of oxygen from the aerobic reactor was given by Eq 3.16 above. Substitution of Eq 3.20 for the molar mass flow rate of the effluent gas $n(AIR)_{out}$ yields:

$$n(O_2)_{out} = n(AIR)_{out} \cdot \frac{\%(O_2)_{out}}{100} = \frac{(100 - \%(O_2)_{in})}{(100 - \%(O_2)_{out} - \%(CO_2)_{out})} \cdot \frac{\%(O_2)_{out}}{100} \quad \text{...kmol/h} \quad (3.21)$$

The rate at which oxygen is **utilised** within the aerobic reactor is calculated from the difference between the influent and effluent oxygen molar mass flow rates. i.e. it is accepted that there is no accumulation of dissolved oxygen in the sludge[2].

$$n(O_2)_{ut} = n(O_2)_{in} - n(O_2)_{out} \quad \text{...kmol/h} \quad (3.22)$$

where:

$n(O_2)_{ut}$ = The molar mass rate of oxygen utilisation (kmol/h)

Substitution of Eq 3.14 for the **influent** molar mass flow rate of oxygen $n(O_2)_{in}$ and Eq 3.21 for the **effluent** molar mass flow rate of oxygen $n(O_2)_{out}$ yields:

$$n(O_2)_{ut} = n(AIR)_{in} \left(\frac{\%(O_2)_{in}}{100} - \frac{\%(O_2)_{out}}{100} \cdot \frac{(100 - \%(O_2)_{in})}{(100 - \%(O_2)_{out} - \%(CO_2)_{out})} \right) \quad \text{...kmol/h} \quad (3.23)$$

3.2.6 The Carbon Dioxide Mass Balance

It is assumed that all the carbon dioxide generated through biological respiration exits the reactor in the effluent gas stream. i.e. a negligible amount remains in solution as dissolved NH_4HCO_3. This aspect was examined by Messenger *et al* (1992) on the Milnerton aerobic reactor and it was found that less than 5% of the carbon dioxide remained in solution. The carbon dioxide mass balance is therefore as follows:

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[2] This was the case for most of the time in the Athlone aerobic reactor. For the instances where this did not apply, identified by detectable DO concentrations in the aerobic reactor, steady state conditions did not prevail and the data were not included in the calculations.

$$n(CO_2)_{gen} = n(CO_2)_{out} = n(AIR)_{out} \cdot \frac{\%(CO_2)_{out}}{100} \quad \text{...kmol/h} \quad (3.24)$$

where:

$n(CO_2)_{gen}$ = The molar mass rate of carbon dioxide generation (kmol/h)

Substitution of Eq 3.20 for $n(AIR)_{out}$ yields:

$$n(CO_2)_{gen} = nCO_2)_{out} = n(AIR)_{in}\left(\frac{\%(CO_2)_{out}}{100} \cdot \frac{(100 - \%(O_2)_{in})}{(100 - \%(O_2)_{out} - \%(CO_2)_{out})}\right) \quad \text{...kmol/h} \quad (3.25)$$

3.2.7 The Respiration Quotient

The respiration quotient Y_{CO2} is defined as the number of moles of carbon dioxide generated per mole of oxygen utilised by the bacteria. i.e.

$$Y_{CO_2} = \frac{n(CO_2)_{gen}}{n(O_2)_{ut}} \quad \text{...kmol/kmol} \quad (3.26)$$

Substitution of Eq 3.25 for $n(CO_2)_{gen}$ and Eq 3.23 for $n(O_2)_{ut}$ yields:

$$Y_{CO_2} = \frac{\%(CO_2)_{out}(100 - \%(O_2)_{in})}{\%(O_2)_{in}(100 - \%(O_2)_{out} - \%(CO_2)_{out}) - \%(O_2)_{out}(100 - \%(O_2)_{in})} \quad \text{...kmol/kmol} \quad (3.27)$$

In this investigation both the $\%(CO_2)_{out}$ and $\%(O_2)_{out}$ volumetric concentrations were measured in the effluent gas stream, enabling the respiration quotient to be determined (see Section 4.3.6). From the 118 paired measurements of $\%(CO_2)_{out}$ and $\%(O_2)_{out}$ the average respiration quotient Y_{CO2} over the investigation period was calculated to be:

$$Y_{CO2} = 0.70 \text{ mol } (CO_2)_{gen}/\text{mol}(O_2)_{ut}$$

This value compares very favourably with the 0.67 value for Y_{CO2} measured by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor, and indicates that reliable measurements were obtained with the combustion analyser. The combustion analyser results were also checked periodically with the ORSAT apparatus (see Messenger *et al*, 1992 for details).

Accepting a constant value for Y_{CO2} allows the $\%(CO_2)_{out}$ term to be eliminated from the equations and replaced by a Y_{CO2}, $\%(O_2)_{out}$ and $\%(O_2)_{in}$ function obtained by rearranging Eq 3.26 in terms of $\%(CO_2)_{out}$: [3]

$$\%(CO_2)_{out} = \frac{100 \times Y_{CO_2}(\%(O_2)_{in} - \%(O_2)_{out})}{(100 + Y_{CO_2}\%(O_2)_{in} - \%(O_2)_{in})} \quad \text{...\% v/v} \quad (3.28)$$

Substitution of Eq 3.28 for $\%(CO_2)_{out}$ into the equations for:

$n(AIR)_{out}$ The molar mass flow rate of the effluent gas stream (Eq 3.20)
$n(O_2)_{out}$ The molar mass flow rate of oxygen in the effluent gas stream (Eq 3.21)
$n(O_2)_{ut}$ The molar mass flow rate of oxygen utilised in the aerobic reactor (Eq 3.23)
$n(CO_2)_{gen}$ The molar mass flow rate of CO_2 generated in the aerobic reactor (Eq3.25)

yields:

$$n(AIR)_{out} = n(AIR)_{in} \cdot \frac{(100 + Y_{CO2}.\%(O_2)_{in} - \%(O_2)_{in})}{(100 + Y_{CO2}.\%(O_2)_{out} - \%(O_2)_{out})} \quad \text{...kmol/h} \quad (3.29)$$

$$n(O_2)_{out} = n(AIR)_{in} \cdot \frac{(100 + Y_{CO2}.\%(O_2)_{in} - \%(O_2)_{in})}{(100 + Y_{CO2}.\%(O_2)_{out} - \%(O_2)_{out})} \cdot \frac{\%(O_2)_{out}}{100} \quad \text{...kmol/h} \quad (3.30)$$

$$n(O_2)_{ut} = n(AIR)_{in} \cdot \frac{(\%(O_2)_{in} - \%(O_2)_{out})}{(100 + Y_{CO2}.\%(O_2)_{out} - \%(O_2)_{out})} \quad \text{...kmol/h} \quad (3.31)$$

$$n(CO_2)_{gen} = Y_{CO2}.n(AIR)_{in} \cdot \frac{(\%(O_2)_{in} - \%(O_2)_{out})}{(100 + Y_{CO2}.\%(O_2)_{out} - \%(O_2)_{out})} \quad \text{...kmol/h} \quad (3.32)$$

Discussion on the Use of a Constant Value for the Respiration Quotient

In the analysis of the oxygen mass and heat balance results, a problem of using a constant Y_{CO2} value of 0.70 for all steady state periods is that it reduces data individuality. The variation in the measured Y_{CO2} values (see Figure 4.4) was however small and it was considered acceptable to take Y_{CO2} as constant. This is acceptable because the effect of Y_{CO2} on the heat balance is negligible even if it were assumed to be unity. The principle reason why Y_{CO2} needed to be known is not for the accuracy of the

[3] The use of this equation (Eq 3.28) does not imply that $\%(CO_2)_{out}$ does not need to be measured. It does, to obtain an accurate measure of the Y_{CO2} value.

heat balance but for the accurate determination of the (mass) oxygen utilisation rate ***OUR*** (Messenger *et al* 1992).

In heat balances on thermophilic aerobic reactors, in particular those using air, sometimes the effluent molar gas flow rate is assumed equal to the influent molar gas flow rate, which assumes that the respiration quotient Y_{CO2} is unity. Assuming this obviates having to measure the effluent gas molar (or volumetric) flow rate, and the oxygen utilisation rate is calculated from only two measurements i.e. the influent gas flow rate and the effluent gas (% $^{v}/_{v}$) oxygen concentration. However, if Y_{CO2} is not unity, as has been found by Messenger *et al* (1992) then this approach leads to considerable error (up to 20%) in the calculated ***OUR***. With an additional measurement i.e. the (% $^{v}/_{v}$) CO_2 in the effluent gas, neither is it necessary to assume unity for Y_{CO2} nor is it necessary to measure the effluent gas volumetric flow rate to obtain an accurate estimation of the oxygen utilisation rate ***OUR***. Instead, the ***OUR*** calculation is based on a more accurate assumption that the molar mass flow rates of the dry influent and effluent equivalent nitrogen (non-oxygen) components are equal.

3.2.8 Gas Component Terms Expressed in Units of Actual Mass Flow Rates

The above expressions for the gas component terms were derived in terms of molar mass flow rates (kmol/h). By definition, molar mass is linked to actual mass via the molecular mass. The molecular mass for oxygen is 32.0, for carbon dioxide is 44.0, and for dry atmospheric air is 29.0 (Mayhew and Rogers, 1977). Accordingly it can be written:

$$M(O_2)_{in,ut,out} = n(O_2)_{in,ut,out} . m_w(O_2) \quad (where\ m_w(O_2) = 32.0) \quad ...kg(O_2)/h \quad (3.33)$$

$$M(CO_2)_{gen,out} = n(CO_2)_{gen,out} . m_w(CO_2) \quad (where\ m_w(CO_2) = 44.0)...kg(CO_2)/h \quad (3.34)$$

$$M(AIR)_{in} = n(AIR)_{in} . m_w(AIR) \quad (where\ m_w(AIR) = 29.0) \quad ...kg(AIR)/h \quad (3.35)$$

where:

$M(O_2)_{in,ut,out}$ = Mass flow rate of oxygen into/utilised/leaving the aerobic reactor ($kg(O_2)/h$)

$M(CO_2)_{gen,out}$ = Mass flow rate of the CO_2 generated/leaving the aerobic reactor ($kg(CO_2)/h$)

$M(AIR)_{in}$ = Dry mass flow rate of the influent gas stream (kg(AIR)/h)

The flow rate of dry air into the aerobic reactor $Q(AIR)_{in}$ was monitored on a volumetric basis and reported as dry air in units of m^3(STP)/h corrected to 20°C and 101.3 kPa (760mmHg). From Mayhew and Rogers (1977) the density of dry atmospheric air at 20°C

and 101.3 Kpa is given as 1.205 kg/m^3 (a value calculated from an application of the ideal gas equation). It follows therefore that:

$$M(AIR)_{in} = Q(AIR)_{in} \cdot \rho(AIR) \qquad (where\ \rho(AIR) = 1.205 kg/m^3) \ldots kg(AIR)/h \quad (3.36)$$

Incorporating the equations linking molecular mass with actual mass (Eq's 3.33-35) and the density equation (Eq 3.36) into the gas component terms yields the following set of equations in terms of actual mass:

$$M(O_2)_{in} = \frac{m_w(O_2) \cdot \rho(AIR) \cdot Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\%(O_2)_{in}}{100} \quad \ldots kg(O_2)/h \quad (3.37)$$

$$M(N_2)_{in,out} = \frac{m_w(N_2) \cdot \rho(AIR) \cdot Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{(100 - \%(O_2)_{in})}{100} \quad \ldots kg(N_2)/h \quad (3.38)$$

$$M(O_2)_{out} = \frac{m_w(O_2) \cdot \rho(AIR) \cdot Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{(100 + Y_{CO2} \cdot \%(O_2)_{in} - \%(O_2)_{in})}{(100 + Y_{CO2} \cdot \%(O_2)_{out} - \%(O_2)_{out})} \cdot \frac{\%(O_2)_{out}}{100}$$
$$\ldots kg(O_2)/h \quad (3.39)$$

$$M(O_2)_{ut} = \frac{m_w(O_2) \cdot \rho(AIR) \cdot Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{(\%(O_2)_{in} - \%(O_2)_{out})}{(100 + Y_{CO2} \cdot \%(O_2)_{out} - \%(O_2)_{out})}$$
$$\ldots kg(O_2)/h \quad (3.40)$$

$$M(CO_2)_{gen} = \frac{Y_{CO2} \cdot m_w(CO_2) \cdot Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{(\%(O_2)_{in} - \%(O_2)_{out})}{(100 + Y_{CO2} \cdot \%(O_2)_{out} - \%(O_2)_{out})}$$
$$\ldots kg(CO_2)/h \quad (3.41)$$

3.2.9 Calculation of the Oxygen Utilisation Rate

By definition (see Section 3.1.5) the oxygen utilisation rate ***OUR*** is obtained by dividing the mass rate of oxygen utilisation in the reactor $M(O_2)_{ut}$ (see Eq 3.40 above) by the process volume V_p i.e.

$$OUR = \frac{m_w(O_2) \cdot \rho(AIR) \cdot Q(AIR)_{in}}{m_w(AIR) \cdot V_p} \cdot \frac{(\%(O_2)_{in} - \%(O_2)_{out})}{(100 + Y_{CO2} \cdot \%(O_2)_{out} - \%(O_2)_{out})}$$
$$\ldots kg(O_2)/m^3.h \quad (3.42)$$

The operational and physical data appropriate for application of the derived equation (Eq 3.42) for the rate of oxygen utilisation ***OUR*** to the Athlone aerobic reactor are contained in Table 3.2 below.

Table 3.2 Operational and Physical Data for Application of the Derived Equation for Oxygen Utilisation Rate in the Athlone Aerobic Reactor

Parameter	Value	Units
$\%(O_2)_{in}$	21.0	% v/v
$m_w(AIR)$	29.0	kg/kmol
$m_w(O_2)$	32.0	kg/kmol
V_p	184.0	m^3
Y_{CO2}	0.70	$mol(CO_2)_{gen}/mol(O_2)_{ut}$
$\rho(AIR)$ at STP	1.205	kg/m^3

Substitution of the values from Table 3.2 into Eq 3.42 yields:

$$OUR = \frac{32 \times 1.205\ Q(AIR)_{in}\left(21.0 - \%(O_2)_{out}\right)}{29 \times 184\left(100 - \%(O_2)_{out} + 0.70 \times \%(O_2)_{out}\right)} \quad \text{...kg}(O_2)/m^3.h \quad (3.43)$$

Which simplifies to:

$$OUR = \frac{Q(AIR)_{in}}{138} \cdot \frac{(21 - \%(O_2)_{out})}{(100 - 0.3 \times \%(O_2)_{out})}$$

Equation 3.44 The Oxygen Utilisation Rate *OUR* ($kg(O_2)/m^3.h$)

To give a visual impression of Eq 3.44, a plot of ***OUR*** versus $\%(O_2)_{out}$ for different influent air flow rates $Q(AIR)_{in}$ (m^3(STP)/h) is given in Figure 3.4 below.

3.2.10 Calculation of the Oxygen Supply Rate

By definition (see Section 3.1.5) the oxygen supply rate ***OSR*** is obtained by dividing the mass flow rate of oxygen into the reactor $M(O_2)_{in}$ (see Eq 3.40 above) by the process volume V_p i.e.

$$OSR = \frac{m_w(O_2) \cdot \rho(AIR) \cdot Q(AIR)_{in}}{m_w(AIR) \cdot V_p} \cdot \frac{\%(O_2)_{in}}{100} \quad \text{...kg}(O_2)/m^3.h \quad (3.45)$$

Substitution of the relevant data from Table 3.2 into Eq 3.45 yields:

$$OSR = \frac{32 \times 1.205 \times Q(AIR)_{in}}{29 \times 184} \cdot \frac{21}{100} \quad \text{...kg}(O_2)/m^3.h \quad (3.46)$$

Which simplifies to:

$$OSR = 0.00151 \times Q(AIR)_{in}$$

Equation 3.47 The Oxygen Supply Rate *OSR* ($kg(O_2)/m^3.h$)

3.2.11 Calculation of the Oxygen Transfer Efficiency

The oxygen transfer efficiency is by definition (see Section 3.1.5):

$$OTE = \frac{OUR}{OSR} \times 100 \quad ...\% \quad (3.48)$$

Substitution of Eq 3.44 for *OUR* and Eq 3.47 for *OSR* yields Eq. 3.49 below. A graphical representation of Eq. 3.41 is given in Figure 3.4.

$$OTE = \frac{(100 - 4.76 \times \%(O_2)_{out})}{(100 - 0.3 \times \%(O_2)_{out})} \times 100$$

Equation 3.49 The Oxygen Transfer Efficiency *OTE* (%)

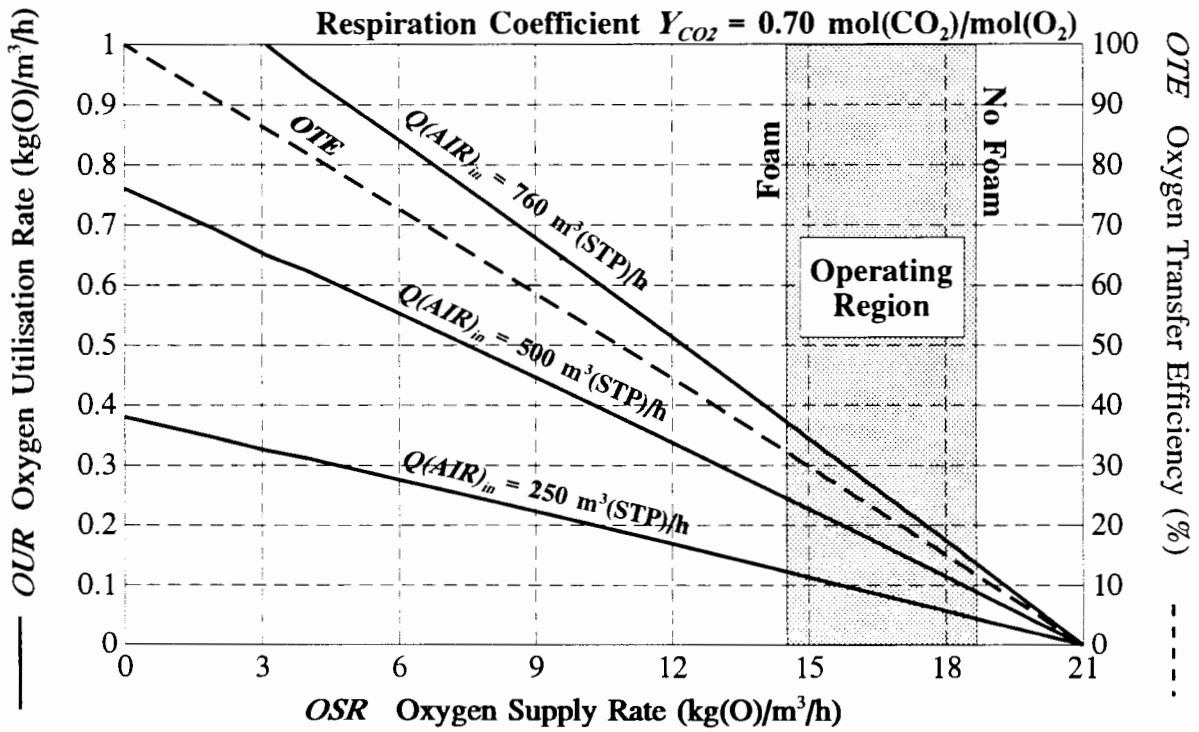

Figure 3.4 Oxygen Utilisation Rate *OUR* (solid lines) and Oxygen Transfer Efficiency *OTE* (dotted lines) versus Percentage Oxygen in the Dry Effluent Gas stream $\%(O_2)_{out}$ ($^v/_v$) for different Influent Air Flow Rates $Q(AIR)_{in}$ (m^3(STP)/h), from Equations 3.44 and 3.49 respectively.

3.3 THE OXYGEN UTILISATION RATE (USING AIR + OXYGEN)

3.3.1 Introduction

Phase II of the Research Programme investigated the feasibility of supplementing the aeration of the aerobic reactor with pure oxygen. Consequently it was necessary to derive expressions to accurately estimate the supply, transfer, and utilisation of oxygen with oxygen supplementation taking place. The derivation is based on the solution of the mass balances of the different components (nitrogen, oxygen, and carbon dioxide) in the combined gas (air and oxygen) stream. The measurable parameters selected for the purpose of solving the gas component mass balances were:

$Q(AIR)_{in}$ = The volumetric air flow rate into the aerobic reactor (m^3(STP)/h)

$M(O_2)_{in}^{O2}$ = The mass flow rate of pure oxygen injected into the sludge recirculation line (kg(O_2)/h)

$\%(O_2)_{out}^{AIR+O2}$ = The oxygen concentration in the dry effluent vent gas during oxygen supplementation ($\%^v/_v$)

$\%(O_2)_{out}^{AIR}$ = The oxygen concentration in the dry effluent vent gas when oxygenation is with air alone ($\%^v/_v$)

In order to solve the gas component mass balance it was necessary to measure the oxygen concentration in the vent gas with oxygen supplementation being both on and off. The procedure for doing this is described in Section 2.4.11 above. A description of the terms employed in the derivation is presented in Table 3.3 below.

Table 3.3 Description of the Terms Used in the Derivation of the Equations Describing the Oxygen Utilisation Rate During Pure Oxygen Supplementation.

Position	Parameter	Description
Main Symbol	*n,M,%*	Denotes molar mass flow rate, actual mass flow rate and percentage volumetric concentration ($^v/_v$).
Bracketed	O_2,N_2,CO_2	Indicates the component gas represented..
Subscript	*in,out,ut,gen*	Signifies whether term relates to the influent or effluent gas, or is connected to the utilisation of oxygen or the generation of carbon dioxide.
Superscript	*AIR*,O_2,*AIR*+O_2	Signifies the air and oxygen streams independently or the combined gas flow during oxygen supplementation

During phase II, the combustion analyser was no longer available and no other suitable analytical equipment was available to measure the carbon dioxide concentration in the vent gas. However, sufficient confidence was placed in the value determined for the respiration coefficient Y_{CO2} during phase I (see Section 4.3.6) for the value Y_{CO2}=0.70 mol(CO_2)/mol(O_2) to be used to solve the carbon dioxide mass balance. The average value for Y_{CO2} of 0.70 compares very favourably with the 0.67 value observed by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor.

In the derivation that follows, it is accepted that the pure oxygen injection system and the diffused air aeration system operate independently of each other; i.e. neither one influences the performance of the other. The validity of this assumption was later confirmed by the accuracy of the heat balance calculated during phase II, plus the fact that the pure oxygen transfer efficiency did not change with changing air flow rate (see Section 7.3.4). The basic philosophy of the assumption is that each system contributes individually to the overall oxygen transfer rate, which under oxygen limiting conditions is equal to the biological utilisation rate. viz:

$$OUR^{AIR+O2} = OUR^{O2} + OUR^{AIR} = OTR^{AIR+O2} = OTR^{O2} + OTR^{AIR} \quad ...kg(O_2)/m^3.h \quad (3.50)$$

3.3.2 The Composition of the Influent and Effluent Gas Streams

During phase II of the Research Programme the aerobic reactor was oxygenated with **atmospheric air** through the course bubble diffusers set in the base of the reactor (see Section 2.3.4) and **pure oxygen** was injected into a Venturi in the sludge recirculation line[4] (see Section 2.3.5).

Influent Gas Stream (Air + Pure Oxygen)

The composition of the influent air stream was as described in Section 3.2.2 above (see Equations 3.13 and 3.14). Adding the molar mass flow of pure oxygen from the injection system, the combined influent gas stream can be characterised by the following equations:

$$n(N_2)_{in}^{AIR} = n(AIR)_{in} \cdot \frac{\%(N_2)_{in}^{AIR}}{100} \quad ...kmol/h \quad (3.51)$$

[4] The sludge recirculation line was modified to incorporate oxygen injection; two larger motors were installed on the mixing pumps and the length of pipework was increased with a narrower bore. The reasons for the alterations are given in Section 2.3.7.

$$n(O_2)_{in}^{AIR} = n(AIR)_{in} \cdot \frac{\%(O_2)_{in}^{AIR}}{100} \quad \text{...kmol/h} \quad (3.52)$$

$$n(O_2)_{in}^{O2} = \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} \qquad (where\ m_w(O_2) = 32) \quad \text{...kmol/h} \quad (3.53)$$

$$n(GAS)_{in}^{AIR+O2} = n(N_2)_{in}^{AIR} + n(O_2)_{in}^{AIR} + n(O_2)_{in}^{O2} \quad \text{...kmol/h} \quad (3.54)$$

where:

$n(N_2,O_2)_{in}^{AIR}$ = The influent molar mass flow rate of nitrogen and oxygen entering the aerobic reactor from the air stream (kmol/h).

$n(O_2)_{in}^{O2}$ = The influent molar mass flow rate of oxygen entering the aerobic reactor from the pure oxygen stream (kmol/h).

$n(GAS)_{in}^{AIR+O2}$ = The combined molar mass gas flow (air + pure oxygen) into the aerobic reactor (kmol/h)

Effluent Gas Stream (Air + Pure Oxygen)

To allow for the oxygenation characteristics of the pure oxygen injection system and the air aeration system to be determined independently the oxygen fraction of the effluent gas stream is considered to consist of two component parts, namely, the oxygen in the air stream $n(O_2)_{out}^{AIR}$ and the oxygen derived from the pure oxygen injection stream $n(O_2)_{out}^{O2}$. Likewise, the carbon dioxide which is generated inside the aerobic reactor as a result of biological respiration is considered to consist of two component parts, namely, that resulting from the utilisation of oxygen derived from the air stream $n(CO_2)_{out}^{AIR}$ and that resulting from the utilisation of the pure oxygen $n(CO_2)_{out}^{O2}$. The combined effluent gas stream can be characterised by the following equations:

$$n(N_2)_{out}^{AIR} = n(N_2)_{out}^{AIR+O2} = n(GAS)_{out}^{AIR+O2} \cdot \%(N_2)_{out}^{AIR+O2} \quad \text{...kmol/h} \quad (3.55)$$

$$n(O_2)_{out}^{AIR+O2} = n(O_2)_{out}^{AIR} + n(O_2)_{out}^{O2} = n(GAS)_{out}^{AIR+O2} \cdot \%(O_2)_{out}^{AIR+O2} \quad \text{...kmol/h} \quad (3.56)$$

$$n(CO_2)_{out}^{AIR+O2} = n(CO_2)_{out}^{AIR} + n(CO_2)_{out}^{O2} = n(GAS)_{out}^{AIR+O2} \cdot \%(CO_2)_{out}^{AIR+O2} \quad \text{...kmol/h} \quad (3.57)$$

$$n(GAS)_{out}^{AIR+O2} = n(N_2)_{out}^{AIR} + n(O_2)_{out}^{AIR} + n(O_2)_{out}^{O2} + n(CO_2)_{out}^{AIR} + n(CO_2)_{out}^{O2} \quad \text{...kmol/h} \quad (3.58)$$

where:

$n(N_2,O_2,CO_2)_{out}^{AIR}$	=	The effluent molar mass flow rate of nitrogen and oxygen originating from the air stream and the effluent molar mass flow rate of carbon dioxide originating from the utilisation of oxygen from the air stream (kmol/h).
$n(O_2,CO_2)_{out}^{O2}$	=	The effluent molar mass flow rate of oxygen originating from the pure oxygen stream and the effluent molar mass flow rate of carbon dioxide originating from the utilisation of the pure oxygen (kmol/h).
$n(N_2,O_2,CO_2)_{out}^{AIR+O2}$	=	The total molar mass flow rate of nitrogen, oxygen, and carbon dioxide respectively leaving the aerobic reactor in the combined (air+pure oxygen) vent gas stream (kmol/h).
$n(GAS)_{out}^{AIR+O2}$	=	The combined molar mass gas flow rate (air + pure oxygen) leaving the aerobic reactor (kmol/h).

A summary of the combined gas stream composition and the mass balances on the different gas components is given in Figure 3.5 below.

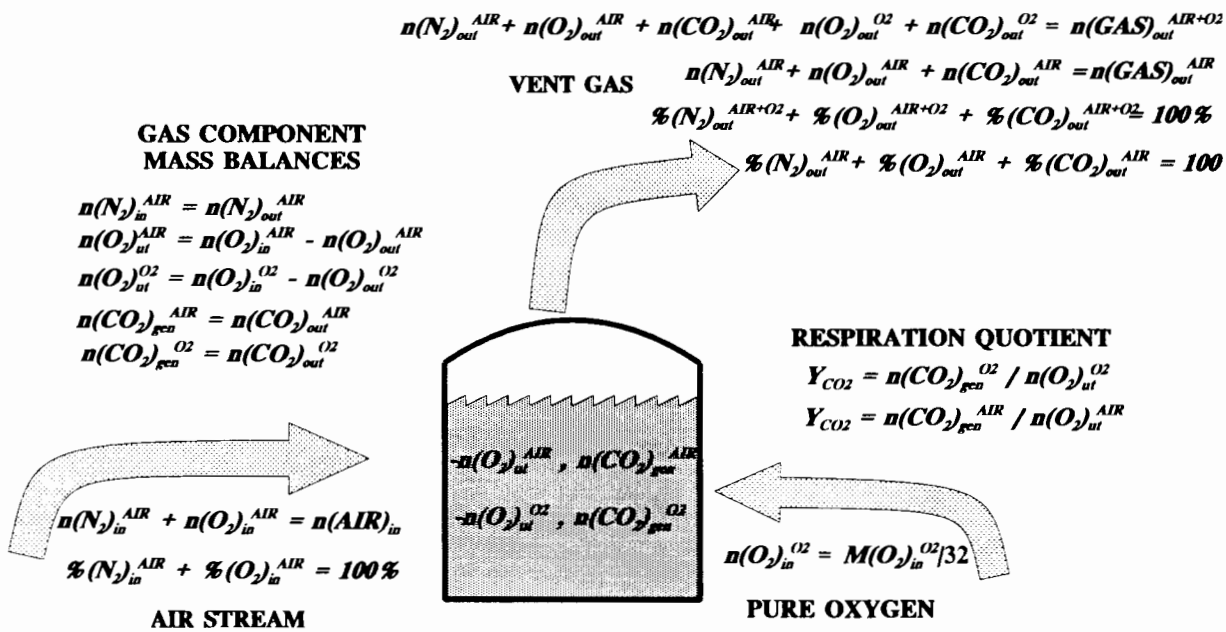

Figure 3.5 Gas Components Across the Aerobic Reactor During Pure Oxygen Supplementation

3.3.3 The Equivalent Nitrogen Gas Mass Balance

With oxygen supplementation taking place, as with operation using air alone, the equivalent nitrogen[5] molar mass flow rate is assumed to remain unchanged through the reactor i.e. there is negligible dissolution of nitrogen and other non-oxygen air components into the sludge.

$$n(N_2)_{in}^{AIR} = n(N_2)_{out}^{AIR} \quad \text{...kmol/h} \quad (3.59)$$

By accepting that the two oxygenation systems operate independently of each other, use can be made of the previously derived equation Eq 3.38 (refer Section 3.2.8 above) expressed in molar mass units. This places the molar mass flow rate of nitrogen $n(N_2)_{in,out}^{AIR}$ in terms of the measurable parameter $Q(AIR)_{in}$, the volumetric flow rate of the air stream into the reactor, viz:

$$n(N_2)_{in}^{AIR} = n(N_2)_{out}^{AIR} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 - \%(O_2)_{in}^{AIR}\right)}{100} \quad \text{.kmol/h} \quad (3.60)$$

3.3.4 The Oxygen Gas Mass Balance

Oxygen enters the aerobic reactor from two sources, with the air stream and through pure oxygen injection. Making use of the previously derived equation Eq 3.37 (refer Section 3.2.8 above), in molar mass units, the flow rate of oxygen entering the reactor with the air stream is given by

$$n(O_2)_{in}^{AIR} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\%(O_2)_{in}^{AIR}}{100} \quad \text{.kmol/h} \quad (3.61)$$

The molar mass flow rate of oxygen entering the reactor through pure oxygen injection, in terms of the measurable parameter $M(O_2)_{in}^{O2}$ (the actual mass flow rate of pure oxygen), was given above by Eq 3.53:

$$n(O_2)_{in}^{O2} = \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} \quad \text{.kmol/h} \quad (3.53)$$

Accordingly, the combined molar mass flow of oxygen entering the aerobic reactor from both sources is given by:

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[5] The equivalent nitrogen fraction incorporates all the gas components in atmospheric air excluding oxygen i.e. nitrogen, argon, carbon dioxide and the rare gases (see Section 3.2.2).

$$n(O_2)_{in}^{AIR+O2} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)}.\frac{\%(O_2)_{in}^{AIR}}{100} + \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} \quad \text{.kmol/h} \quad (3.62)$$

The molar mass flow rate of oxygen leaving the reactor, accepting that the oxygen fraction of the vent gas consists of two components (derived from the air and pure oxygen streams separately), was given by Eq 3.56 above as:

$$n(O_2)_{out}^{AIR+O2} = n(O_2)_{out}^{AIR} + n(O_2)_{out}^{O2} = n(GAS)_{out}^{AIR+O2}.\frac{\%(O_2)_{out}^{AIR+O2}}{100} \quad \text{.kmol/h} \quad (3.56)$$

The effluent molar mass flow rate of oxygen, originating from the air stream, $n(O_2)_{out}^{AIR}$ can be determined by switching off the pure oxygen supply rate, monitoring the concentration of oxygen in the vent gas $\%(O_2)_{out}^{AIR}$, and using Eq 3.39 (refer Section 3.2.8 above) expressed in molar mass units viz:

$$n(O_2)_{out}^{AIR} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)}.\frac{\left(100+Y_{CO2}.\%(O_2)_{in}^{AIR}-\%(O_2)_{in}^{AIR}\right)}{\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR}-\%(O_2)_{out}^{AIR}\right)}.\frac{\%(O_2)_{out}^{AIR}}{100} \quad \text{.kmol/h} \quad (3.63)$$

Substitution of Eq 3.63 into Eq 3.56 and rearranging in terms of $n(O_2)_{out}^{O2}$, the effluent molar mass flow rate of oxygen, originating from the pure oxygen stream, is given by:

$$n(O_2)_{out}^{O2} = n(GAS)_{out}^{AIR+O2}.\frac{\%(O_2)_{out}^{AIR+O2}}{100} - \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)}.\frac{\left(100+Y_{CO2}.\%(O_2)_{in}^{AIR}-\%(O_2)_{in}^{AIR}\right)}{\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR}-\%(O_2)_{out}^{AIR}\right)}.\frac{\%(O_2)_{out}^{AIR}}{100} \quad \text{..kmol/h} \quad (3.64)$$

The rate at which pure oxygen is utilised within the aerobic reactor is calculated from the difference between the influent pure oxygen flow rate and the effluent flow rate of (formerly pure) oxygen from the reactor viz:

$$n(O_2)_{ut}^{O2} = n(O_2)_{in}^{O2} - n(O_2)_{out}^{O2} \quad \text{...kmol/h} \quad (3.65)$$

Substitution of Eq 3.53 for the influent molar mass flow rate of pure oxygen $n(O_2)_{in}^{O2}$ and Eq 3.61 for the effluent mass flow rate of oxygen $n(O_2)_{out}^{O2}$ yields:

$$n(O_2)_{ut}^{O2} = \frac{M(O_2)_{in}^{AIR+O2}}{m_w(O_2)} - n(GAS)_{out}^{AIR+O2}.\frac{\%(O_2)_{out}^{AIR+O2}}{100} + \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)}.\frac{\left(100+Y_{CO2}.\%(O_2)_{in}^{AIR}-\%(O_2)_{in}^{AIR}\right)}{\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR}-\%(O_2)_{out}^{AIR}\right)}.\frac{\%(O_2)_{out}^{AIR}}{100} \quad \text{..kmol/h} \quad (3.66)$$

The rate at which the oxygen originating from the air stream is utilised can be determined by switching off the pure oxygen supply rate, monitoring the concentration of oxygen in the vent gas $\%(O_2)_{out}^{AIR}$, and using Eq 3.40 (refer Section 3.2.8 above) expressed in molar mass units viz:

$$n(O_2)_{ut}^{AIR} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\left(\%(O_2)_{in}^{AIR} - \%(O_2)_{out}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} \quad \text{.kmol/h} \quad (3.67)$$

The overall oxygen utilisation rate can be determined by adding the individual rate equations; Eq 3.66 for utilisation by formerly pure oxygen and Eq 3.67 for utilisation by oxygen from the air stream viz:

$$n(O_2)_{ut}^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} - n(GAS)_{out}^{AIR+O2} . \frac{\%(O_2)_{out}^{AIR+O2}}{100} +$$

$$\frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} . \frac{\%(O_2)_{out}^{AIR}}{100} +$$

$$\frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\left(\%(O_2)_{in}^{AIR} - \%(O_2)_{out}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)}$$

...kmol/h (3.68)

which simplifies to:

$$n(O_2)_{ut}^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} - n(GAS)_{out}^{AIR+O2} . \frac{\%(O_2)_{out}^{AIR+O2}}{100} + \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\%(O_2)_{in}^{AIR}}{100}$$

...kmol/h (3.69)

3.3.5 The Carbon Dioxide Gas Mass Balance

With oxygen supplementation taking place, as with operation using air alone, all the carbon dioxide generated through biological respiration is assumed to exit the reactor in the effluent gas stream. i.e. a negligible amount remains in solution as dissolved NH_4HCO_3. As in describing the fate of the influent oxygen previously, the carbon dioxide generated is defined in terms of the origin of the oxygen utilised during biological respiration. i.e. air stream or pure oxygen supply. Accordingly:

$$n(CO_2)_{out}^{AIR} = n(CO_2)_{gen}^{AIR} \quad \text{...kmol/h} \quad (3.70)$$

$$n(CO_2)_{out}^{O2} = n(CO_2)_{gen}^{O2} \quad \text{...kmol/h} \quad (3.71)$$

The molar mass flow rates of carbon dioxide from the aerobic reactor can be expressed in terms of the rates of oxygen utilisation of the two oxygen fractions (air stream + pure oxygen) via the respiration quotient Y_{CO2}. A value of Y_{CO2} = 0.70 kmol$(CO_2)_{gen}$/kmol$(O_2)_{ut}$ was determined during Phase I of the investigation (see Section 4.3.6). Sufficient confidence was placed in the accuracy of this value for it to be used to resolve the carbon dioxide mass balance during phase II. Accordingly, it was not necessary to measure the effluent gas carbon dioxide concentration[6]. It follows therefore:

$$n(CO_2)_{out}^{AIR} = n(CO_2)_{gen}^{AIR} = Y_{CO2}.n(O_2)_{ut}^{AIR} \quad \text{...kmol/h} \quad (3.72)$$

$$n(CO_2)_{out}^{O2} = n(CO_2)_{gen}^{O2} = Y_{CO2}.n(O_2)_{ut}^{O2} \quad \text{...kmol/h} \quad (3.73)$$

Substitution of Eq 3.67 above for $n(O_2)_{ut}^{AIR}$ in to Eq 3.71 yields:

$$n(CO_2)_{out}^{AIR} = \frac{Y_{CO2}.\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\left(\%(O_2)_{in}^{AIR} - \%(O_2)_{out}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} \quad \text{...kmol/h} \quad (3.74)$$

Substitution of Eq 3.66 above for $n(O_2)_{ut}^{O2}$ in to Eq 3.72 yields:

$$n(CO_2)_{out}^{O2} = \frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} - Y_{CO2}.n(GAS)_{out}^{AIR+O2} . \frac{\%(O_2)_{out}^{AIR+O2}}{100} + \frac{Y_{CO2}.\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} . \frac{\%(O_2)_{out}^{AIR}}{100} \quad \text{...kmol/h} \quad (3.75)$$

The combined molar mass flow rate of carbon dioxide leaving the aerobic reactor is obtained by summing the two component parts as follows:

$$n(CO_2)_{out} = n(CO_2)_{out}^{AIR} + n(CO_2)_{out}^{O2} \quad \text{...kmol/h} \quad (3.76)$$

Substitution of Eq 3.74 and Eq 3.75 yields:

$$n(CO_2)_{out}^{AIR+O2} = \frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} - Y_{CO2}.n(GAS)_{out}^{AIR+O2} . \frac{\%(O_2)_{out}^{AIR+O2}}{100} + \frac{Y_{CO2}.\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\%(O_2)_{in}^{AIR}}{100} \quad \text{...kmol/h} \quad (3.77)$$

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[6] During Phase II, no suitable instrument was available to measure the carbon dioxide concentration in the effluent gas stream.

3.3.6 The Effluent Molar Mass Flow Rate from The Aerobic Reactor

The only unknown term in the equations derived thus far to characterise the oxygenation characteristics of the system when using air with pure oxygen supplementation is the overall molar mass flow rate of the effluent gas stream $n(GAS)_{out}^{AIR+O2}$. The overall flow rate was defined in Section 3.3.2 above in terms of the individual component gases viz:

$$n(GAS)_{out}^{AIR+O2} = n(N_2)_{out}^{AIR} + n(O_2)_{out}^{AIR} + n(O_2)_{out}^{O2} + n(CO_2)_{out}^{AIR} + n(CO_2)_{out}^{O2} \quad \text{.kmol/h} \quad (3.58)$$

EquationNo: 3.60 3.63 3.64 3.73 3.74

Substitution of the Equations indicated for the individual component gases yields:

$$n(GAS)_{out}^{AIR+O2} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{(100 - \%(O_2)_{in})}{70} +$$

$$\frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} \cdot \frac{\%(O_2)_{out}^{AIR}}{100} +$$

$$n(GAS)_{out}^{AIR+O2} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{100} - \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} \cdot \frac{\%(O_2)_{out}^{AIR}}{100}$$

$$\frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} - Y_{CO2}.n(GAS)_{out}^{AIR+O2} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{100} +$$

$$\frac{Y_{CO2}.\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} \cdot \frac{\%(O_2)_{out}^{AIR}}{100} +$$

$$\frac{Y_{CO2}.\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(\%(O_2)_{in}^{AIR} - \%(O_2)_{out}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)}$$

...kmol/h (3.78)

This then simplifies to:

$$n(GAS)_{out}^{AIR+O2} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 - \%(O_2)_{in}^{AIR}\right)}{100} + n(GAS)_{out}^{AIR+O2} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{100} +$$

$$\frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} - Y_{CO2}.n(GAS)_{out}^{AIR+O2} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{100} + \frac{Y_{CO2}.\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\%(O_2)_{in}^{AIR}}{100}$$

...kmol/h (3.79)

The overall mass flow rate $n(GAS)_{out}^{AIR+O2}$ appears on both sides of Eq 3.77. Rearranging in terms of $n(GAS)_{out}^{AIR+O2}$ yields:

$$n(GAS)_{out}^{AIR+O2} = \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} +$$

$$\frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} \cdot \frac{100}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)}$$

...kmol/h (3.80)

Substitution of Eq 3.80 for $n(GAS)_{out}^{AIR+O2}$ into Eq 3.66, the rate equation for the utilisation of pure oxygen $n(O_2)_{ut}^{O2}$ yields:

$$n(O_2)_{ut}^{O2} = \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} - \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{100}$$

$$\frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} +$$

$$\frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} \cdot \frac{\%(O_2)_{out}^{AIR}}{100}$$

...kmol/h (3.81)

Likewise, substitution of Eq 3.80 for $n(GAS)_{out}^{AIR+O2}$ into Eq 3.66, the overall rate equation for the utilisation of oxygen from both the air and pure oxygen streams $n(O_2)_{ut}^{AIR+O2}$ yields:

$$n(O_2)_{ut}^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{m_w(O_2)} - \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR-O2} - \%(O_2)_{out}^{AIR-O2}\right)} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{100}$$

$$- \frac{Y_{CO2}.M(O_2)_{in}^{O2}}{m_w(O_2)} \cdot \frac{\%(O_2)_{out}^{AIR+O2}}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} + \frac{\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} \cdot \frac{\%(O_2)_{in}^{AIR}}{100}$$

...kmol/h (3.82)

3.3.7 Gas Component Terms in Units of Actual Mass Rates

The above expressions for the oxygen utilisation rate for both oxygen components in the gas stream are derived in terms of molar mass units (kmol/h). By definition, molar mass is linked to actual mass via the molecular mass (see Section 3.2.8 above) viz:

$$M(O_2)_{in,out,ut}^{AIR,O2,-} = n(O_2)_{in,out,ut}^{AIR,O2,-} . m_w(O_2) \qquad (where\ m_w(O_2) = 32.0) \qquad ...kg(O_2)/h \quad (3.83)$$

where:

$M(O_2)_{in,out,ut}^{AIR,O2,-}$ = Mass flow rate of oxygen (from the air stream, pure oxygen stream and combined) into, utilised, and leaving the aerobic reactor ($kg(O_2)/h$)

The mass rate of oxygen utilisation from the oxygen derived from the pure oxygen stream $M(O_2)_{ut}^{O2}$ is therefore as follows:

$$M(O_2)_{ut}^{O2} = M(O_2)_{in}^{O2} - \frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}.\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{m_w(AIR).\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} . \frac{\%(O_2)_{out}^{AIR+O2}}{100} -$$

$$\frac{Y_{CO2}.M(O_2)_{in}^{O2}.\%(O_2)_{out}^{AIR+O2}}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} +$$

$$\frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}.\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{m_w(AIR).\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR} - \%(O_2)_{out}^{AIR}\right)} . \frac{\%(O_2)_{out}^{AIR}}{100}$$

...$kg(O_2)/h$ (3.84)

The overall mass rate of oxygen utilisation $M(O_2)_{ut}^{AIR+O2}$ is given by:

$$M(O_2)_{ut}^{AIR+O2} = M(O_2)_{in}^{O2} - \frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}.\left(100 + Y_{CO2}.\%(O_2)_{in}^{AIR} - \%(O_2)_{in}^{AIR}\right)}{m_w(AIR).\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} . \frac{\%(O_2)_{out}^{AIR+O2}}{100}$$

$$- \frac{Y_{CO2}.M(O_2)_{in}^{O2}.\%(O_2)_{out}^{AIR+O2}}{\left(100 + Y_{CO2}.\%(O_2)_{out}^{AIR+O2} - \%(O_2)_{out}^{AIR+O2}\right)} + \frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}}{m_w(AIR)} . \frac{\%(O_2)_{in}^{AIR}}{100}$$

...$kg(O_2)/h$ (3.85)

3.3.8 Calculation of the Oxygen Utilisation Rate: (Pure Oxygen)

By definition (see Section 3.1.5) the oxygen utilisation rate OUR^{O2} is obtained by dividing the mass rate of oxygen utilisation in the reactor $M(O_2)_{ut}^{O2}$ (see Eq 3.84 above) by the process volume V_p i.e.

$$OUR^{O2} = \frac{M(O_2)_{in}^{O2}}{V_p} - \frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}.\left(100+Y_{CO2}.\%(O_2)_{in}^{AIR}-\%(O_2)_{in}^{AIR}\right)}{V_p.m_w(AIR).\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR+O2}-\%(O_2)_{out}^{AIR+O2}\right)}.\frac{\%(O_2)_{out}^{AIR+O2}}{100}$$

$$-\frac{Y_{CO2}.M(O_2)_{in}^{O2}.\%(O_2)_{out}^{AIR+O2}}{V_p.\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR+O2}-\%(O_2)_{out}^{AIR+O2}\right)}+$$

$$\frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}.\left(100+Y_{CO2}.\%(O_2)_{in}^{AIR}-\%(O_2)_{in}^{AIR}\right)}{V_p.m_w(AIR).\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR}-\%(O_2)_{out}^{AIR}\right)}.\frac{\%(O_2)_{out}^{AIR}}{100}$$

...$kg(O_2)/m^3.h$ (3.86)

Table 3.2 above contains the operational and physical data for application to the derived equation Eq 3.86 for the oxygen utilisation rate OUR^{O2} attributable to the pure oxygen injected into the sludge side stream. Substitution of the values from Table 3.2 into Eq 3.86 yields:

$$OUR^{O2} = \frac{M(O_2)_{in}^{O2}}{184} - \frac{32\times1.205\times.Q(AIR)_{in}.(100+0.70\times21.0-21.0)}{184\times29.0\left(100+0.70\times\%(O_2)_{out}^{AIR+O2}-\%(O_2)_{out}^{AIR+O2}\right)}.\frac{\%(O_2)_{out}^{AIR+O2}}{100} -$$

$$\frac{0.70\times M(O_2)_{in}^{O2}.\%(O_2)_{out}^{AIR+O2}}{184\left(100+0.70\times\%(O_2)_{out}^{AIR+O2}-\%(O_2)_{out}^{AIR+O2}\right)}+\frac{32\times1.205Q(AIR)_{in}(100+0.70\times21.0-21.0)}{184\times29.0\left(100+0.70\%(O_2)_{out}^{AIR}-\%(O_2)_{out}^{AIR}\right)}.\frac{\%(O_2)_{out}^{AIR}}{100}$$

...$kg(O_2)/m^3.h$ (3.87)

which simplifies to:

$$OUR^{O2} = \frac{M(O_2)_{in}^{O2}}{184} - \frac{\%(O_2)_{out}^{AIR+O2}\left(1.25Q(AIR)_{in}+0.7M(O_2)_{in}^{O2}\right)}{184\left(100-0.3\%(O_2)_{out}^{AIR+O2}\right)} + \frac{1.25\%(O_2)_{out}^{AIR}.Q(AIR)_{in}}{184\left(100-0.3\%(O_2)_{out}^{AIR}\right)}$$

Equation 3.88 The "Pure" Oxygen Utilisation Rate OUR^{O2} ($kg(O_2)/m^3.h$)

To give a visual impression of Eq 3.88, a plot of OUR^{O2} versus $\%(O_2)_{out}^{AIR+O2}$ for different influent pure oxygen flow rates $M(O_2)_{in}^{O2}$ ($kg(O_2)/h$) is given in Figure 3.6 below: To enable the equation to be resolved, the influent air flow rate is taken as constant at 760 m^3(STP)/h and the oxygen utilisation rate attributable to the oxygen from the air stream (OUR_{AIR}) is accepted to be constant at 0.15 $kg(O^2)/m^3.h$ with the resulting oxygen concentration in the vent gas $\%(O_2)_{out}^{AIR}$ constant at 18.5% (pure oxygen off).

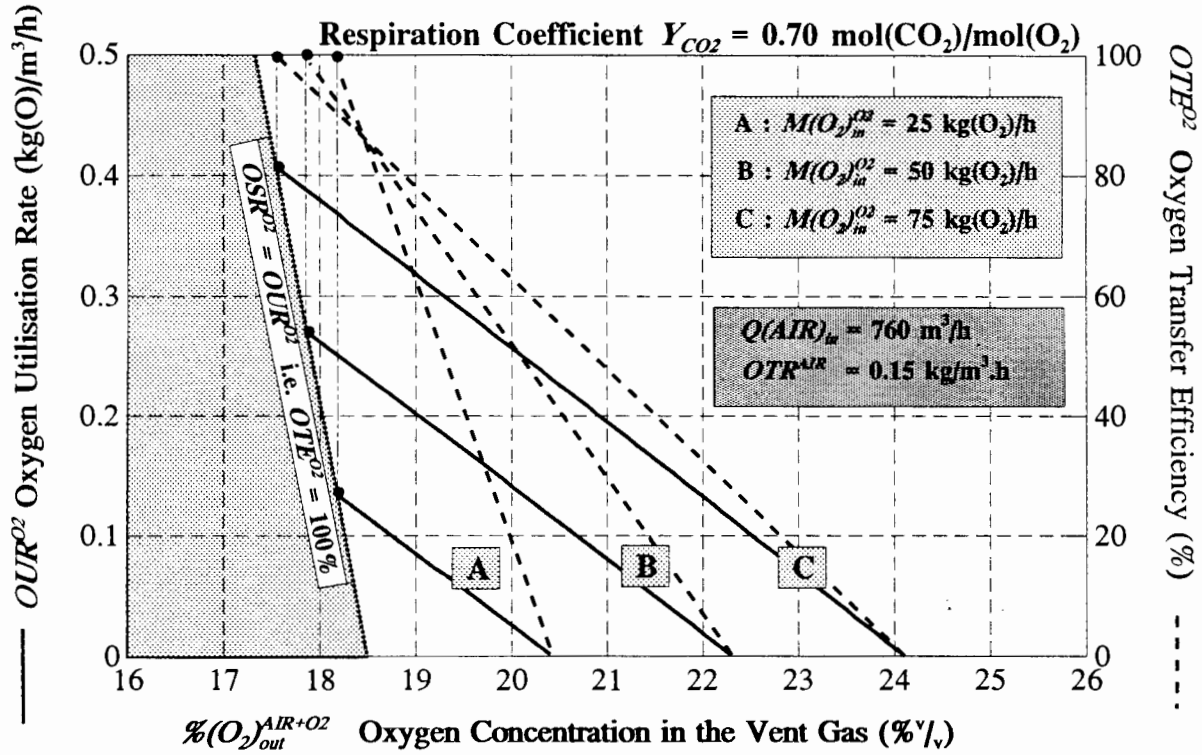

Figure 3.6 Oxygen Utilisation Rate OUR^{O2} (solid lines) and Oxygen Transfer Efficiency OTE^{O2} (dashed lines) versus Percentage Oxygen in the Dry Effluent Gas Stream $\%(O_2)_{out}^{AIR+O2}$ for Different Pure Oxygen Supply Rates $M(O_2)_{in}^{O2}$ (A, B and C) at a Constant Air Flow Rate of 760 m³(STP)/h and an air OTR^{AIR} = 0.15kg(O)/m³.h.

3.3.9 Calculation of the Overall Oxygen Utilisation Rate

The derived expression for the overall mass rate of oxygen utilisation $M(O_2)_{ut}^{AIR+O2}$ was given above by Eq 3.85. The volume specific overall oxygen utilisation rate OUR^{AIR+O2} can be obtained by dividing by the process volume V_p viz:

$$OUR^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{V_p} - \frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}.\left(100+Y_{CO2}.\%(O_2)_{in}^{AIR}-\%(O_2)_{in}^{AIR}\right)}{V_p.m_w(AIR).\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR+O2}-\%(O_2)_{out}^{AIR+O2}\right)}.\frac{\%(O_2)_{out}^{AIR+O2}}{100}$$

$$- \frac{Y_{CO2}.M(O_2)_{in}^{O2}.\%(O_2)_{out}^{AIR+O2}}{V_p.\left(100+Y_{CO2}.\%(O_2)_{out}^{AIR+O2}-\%(O_2)_{out}^{AIR+O2}\right)} + \frac{m_w(O_2).\rho(AIR).Q(AIR)_{in}}{V_p.m_w(AIR)}.\frac{\%(O_2)_{in}^{AIR}}{100}$$

...kg(O_2)/m³.h (3.89)

Substitution of the operational and physical data contained in Table 3.2 above yields:

$$OUR^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{184} - \frac{32\times1.205\times Q(AIR)_{in}.(100+0.70\times21.0-21.0)}{184\times29.0\left(100+0.70\times\%(O_2)_{out}^{AIR-O2}-\%(O_2)_{out}^{AIR-O2}\right)}.\frac{\%(O_2)_{out}^{AIR-O2}}{100}$$

$$-\frac{0.70\times M(O_2)_{in}^{O2}.\%(O_2)_{out}^{AIR-O2}}{184\left(100+0.70\times\%(O_2)_{out}^{AIR-O2}-\%(O_2)_{out}^{AIR+O2}\right)} + \frac{32\times1.205\,Q(AIR)_{in}}{184\times29.0}.\frac{21.0}{100}$$

...$kg(O_2)/m^3.h$ (3.90)

which simplifies to:

$$OUR^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{184} + \frac{0.279\,Q(AIR)_{in}}{184} - \frac{\%(O_2)_{out}^{AIR+O2}\left(1.25Q(AIR)_{in}+0.7M(O_2)_{in}^{O2}\right)}{184\left(100-0.3\%(O_2)_{out}^{AIR+O2}\right)}$$

Equation 3.91 The Overall Oxygen Utilisation Rate OUR^{AIR+O2} ($kg(O_2)/m^3.h$)

3.3.10 Calculation of the Overall Oxygen Supply Rate

The expression for the oxygen supply rate delivered by the air stream (OSR^{AIR}) was given by Eq 3.47 above, viz.

$$OSR^{AIR} = 0.00151 \times Q(AIR)_{in} = \frac{0.279 \times Q(AIR)_{in}}{184} \quad ...kg(O_2)/m^3.h \quad (3.47)$$

The oxygen supply rate delivered by pure oxygen injection (OSR^{O2}) is given by:

$$OSR^{O2} = \frac{M(O_2)_{in}^{O2}}{184} \quad ...kg(O_2)/m^3.h \quad (3.92)$$

The overall oxygen supply rate (OSR^{AIR+O2}) is therefore given by:

$$OSR^{AIR+O2} = \frac{0.279\times Q(AIR)_{in} + M(O_2)_{in}^{O2}}{184}$$

Equation 3.93 The Overall Oxygen Supply Rate OSR^{AIR+O2} ($kg(O_2)/m^3.h$)

3.3.11 Calculation of the Oxygen Transfer Efficiency: Pure Oxygen

By definition the "pure" oxygen transfer efficiency (OTE^{O2}) is given by:

$$OTE^{O2} = \frac{OUR^{O2}}{OSR^{O2}} \times 100 \quad ...\% \quad (3.94)$$

SSubstitution of Eq 3.88 for OUR^{O2} and Eq 3.92 for OSR^{O2} yields:

$$OTE^{O2} = 100 - \frac{\%(O_2)_{out}^{AIR+O2}\left(125Q(AIR)_{in} + 70M(O_2)_{in}^{O2}\right)}{M(O_2)_{in}^{O2}.\left(100 - 0.3\%(O_2)_{out}^{AIR+O2}\right)} + \frac{125\%(O_2)_{out}^{AIR}.Q(AIR)_{in}}{M(O_2)_{in}^{O2}.\left(100 - 0.3\%(O_2)_{out}^{AIR}\right)}$$

Equation 3.95 The "Pure" Oxygen Transfer Efficiency OTE^{O2} (%)

A graphical representation of Eq 3.95 (OTE^{O2} versus $\%(O_2)_{out}^{AIR+O2}$) for different influent pure oxygen flow rates $M(O_2)_{in}^{O2}$ (kg(O_2)/h) is given in Figure 3.6 above. The influent air flow rate is taken as constant at 760 m^3(STP)/h with the oxygen transfer rate attributable to the oxygen from the air stream (OTR^{AIR}) accepted to be constant at 0.15 kg(O^2)/m^3.h with the resultant oxygen concentration in the vent gas $\%(O_2)_{out}^{AIR}$ at 18.5% (pure oxygen off).

3.4 THE AEROBIC REACTOR STEADY STATE HEAT BALANCE

3.4.1 Description of the Heat Terms in the Steady State Heat Balance

The aerobic reactor steady state heat balance was given above in Eq 3.8 which equates the heat sources to the heat losses, the former being the sum of the biological (H_b) and mechanical (H_m) heat input rates, and the latter being the sum of the effluent sludge (H_s) and vent gas (H_g) sensible heat loss rates, the vent gas water vapour (H_v) and wall (H_w) heat loss rates.

Biological heat is generated in the aerobic reactor due to the inefficiency of the aerobic bacteria in synthesising new cell material (see Section 3.1.2). The rate of biological heat generation H_b (MJ/h) is directly linked to the biological oxygen utilisation rate $\mathbf{M}(O_2)_{ut}$ (kg(O_2)/h) via the specific heat yield coefficient Y_h (MJ/kg(O_2)) (see Eq. 3.8).

Mechanical heat is generated from the action of the centrifugal mixing pumps. The impact of the impellers on the sludge imparts kinetic energy to the sludge. This kinetic energy is then transferred to thermal energy through collisions which take place between sludge particles.

Sludge sensible heat is lost when hot sludge is transferred out of the aerobic reactor to the anaerobic digester. The rate of heat loss H_s is dependent on the sludge flow, the heat capacity of the sludge (C_p) and the temperature difference between the effluent (or reactor) and influent sludges $(T(SL)r - T(SL)_{in})$.

As the gas stream passes upwards through the aerobic reactor both heat and water vapour are transferred from the sludge to the gas. Providing there is sufficient contact time, equilibrium will be reached both in terms of heat and mass transfer. Heat will flow from the hot sludge to the cooler air until the two phases are at the same temperature. Sensible heat H_s is lost from the system when the gas stream is vented. The rate of heat loss is dependent on the effluent gas flow rate, the heat capacity of the gas stream (C_p) and the difference between the effluent and influent gas stream temperatures $(T(AIR)_{out} - T(AIR)_{in})$. Under non-foaming conditions it was found that the effluent (vent) gas temperature was generally about 3°C lower than the liquid sludge temperature due to the heat losses which take place in the reactor head space.

The gas stream through the aerobic reactor acquires water vapour from the sludge until the partial pressure of the vapour in the gas stream equals the vapour pressure of the sludge, at which point the gas is considered to be saturated. When the water is vaporised there is a transfer of heat from the sludge to the gas stream and this heat H_v is subsequently lost from the system when the gas stream is vented. The rate of heat loss is dependent on the latent heat of water vaporisation $L_v^{50^\circ C}$ and the mass rate of water vaporisation from the sludge $M(H_2O)_{vap}$.

The heat loss through the walls of the aerobic reactor is due to a combination of conduction, convection and radiation.

Each of the different heat source and loss rate terms is discussed and derived in detail below. A summary of the derived formulae is given in Section 3.9.

3.5 THE VAPOUR HEAT LOSS IN THE EFFLUENT GAS

3.5.1 Introduction

As the gas stream passes upwards through the aerobic reactor, water vapour is transferred from the sludge liquid to the gas stream. Providing there is sufficient contact time, the transfer of water vapour will continue until the partial pressure of the vapour in the gas stream equals the vapour pressure of the sludge liquid, at which point the gas stream is considered to be saturated. When the water is vaporised there is a transfer of heat from the sludge to the gas stream and this heat H_v is lost from the system when the gas stream is vented.

The vapour heat loss in the effluent gas stream H_v is one of two major heat loss terms in the heat balance Eq 3.8 (the other is the sensible heat of the effluent sludge liquid H_s). The relatively high air flow rates which are encountered in the aerobic reactor effect a high rate of water vaporisation from the sludge. Indeed if the effluent air is saturated when the gas is vented from the reactor as much as 5% to 7% of the water flow is vaporised (depending on the retention time) resulting in about one third of the total heat sources being lost via water vaporisation.

In order to quantify the vapour heat loss rate in the effluent gas H_v, consideration needs to be given to (1) the latent heat of vaporisation of water at the temperatures encountered in the effluent gas stream and (2) the water mass loss rate due to vaporisation which takes place within the aerobic reactor. The rate of water vaporisation is calculated from the product of the dry air mass flow rate and the difference between the humidity of the effluent and influent gas streams. Humidity is expressed in terms of the water vapour partial pressure, a parameter which is dependent upon gas stream temperature (a measurable parameter). For the water vapour (and gas sensible) heat loss rate, it is permissible to assume that the influent and effluent dry air mass flow rates are equal (see Section 3.5.3 below) with negligible loss of accuracy. The influent dry air mass flow rate is computed from the measured volumetric flow rate via the air density as outlined above in Eq 3.36.

During phase II, pure oxygen was injected into the sludge recirculation line in addition to the normal air flow through the reactor. The influent mass flow rate of pure oxygen was directly measured by a rotameter flow meter. Allowance is therefore made to accommodate the increase in gas flow rate through the system and hence the higher vapour heat loss rate H_v^{AIR+O2} which results (see Section 3.5.9 below).

3.5.2 The Latent Heat of Vaporisation at Reactor Temperatures

When a pure substance is vaporised from the liquid state at constant pressure, there is no change in temperature but there is a definite transfer of heat from the surroundings to the substance. This effect is commonly called the latent heat of vaporisation, and is defined quantitatively as the heat required to convert one unit mass of a given substance from the liquid state to the gaseous state at a given temperature and pressure.

The latent heat of vaporisation decreases with both increasing temperature and pressure; however where the temperature does not vary over wide limits, it may be assumed that the heat of vaporisation remains constant. The variation in the temperature range encountered in the aerobic reactor is sufficiently small (2.39→2.34MJ/kg in the temperature range 45°→65°C) to make this assumption and a constant value for the heat of vaporisation can be accepted with negligible loss of accuracy. The value for the latent heat of vapourisation at 50°C is selected as this was the most common reactor temperature during phase I, when gas flow rates were at their highest. From Steam Tables by Mayhew and Rogers (1977) the Latent Heat of Vaporisation of tap water (ie low salinity) at 50°C and an external pressure of 1 atm. (101.3 Kpa or 760 mmHg) is given as:

$$L_v^{50°C} = 2.382 \quad \text{...MJ/kg} \quad (3.95)$$

Equation Eq 3.95 describes the heat lost through the vaporisation of 1 kg of water (in the temperature range 45 to 65°C) from the aerobic reactor. The total amount of heat lost from the system due to water vaporisation can therefore be determined if the total mass of water vaporised, termed the water vaporisation rate, can be quantified:

$$H_v = L_v^{50°C} . M(H_2O)_{vap} \quad \text{...MJ/h} \quad (3.96)$$

where:

H_v = Total vapour heat loss from the reactor (MJ/h)

$M(H_2O)_{vap}$ = Water vaporisation rate (kg/h)

3.5.3 The Water Vaporisation Rate

The water vaporisation rate can be equated to the difference between the mass flow rate of water vapour in the effluent and influent gas stream. viz:

$$M(H_2O)_{vap} = M(H_2O)_{in} - M(H_2O)_{out} \quad \text{...kg}(H_2O)\text{/h} \quad (3.97)$$

where:

$M(H_2O)_{out}$ = Mass flow rate of water vapour in the *effluent* gas stream (kg(H_2O)/h)

$M(H_2O)_{in}$ = Mass flow rate of water vapour in the *influent* gas stream (kg(H_2O)/h)

3.5.4 Gas Stream Humidity

It is convenient at this stage to introduce the concept of humidity U. The humidity of a gas is generally defined as the mass of water vapour per unit mass of dry air. For the influent and effluent gas streams the humidity can be expressed as follows:

$$U_{out} = \frac{M(H_2O)_{out}}{M(AIR)_{out}} \quad \ldots (3.98)$$

and

$$U_{in} = \frac{M(H_2O)_{in}}{M(AIR)_{in}} \quad \ldots (3.99)$$

where:

$U_{in,out}$ = Humidity in the influent and effluent gas streams respectively

$M(AIR)_{in,out}$ = Mass flow rate of dry air in the influent and effluent gas streams (kg/h)

In determining the water vaporisation rate, it is accepted that **the mass flow rate of dry air remains unchanged through the system**. Note this assumption is only made for the heat loss terms; it was not made in determining the oxygen utilisation rate (see Section 3.2.3 above). For the air oxygenated reactor this assumption is reasonable for the heat balance and based on the fact that (1) 79% of the influent gas stream (AIR) consists of equivalent nitrogen gas which passes through the process unchanged and (2) the mass of carbon dioxide generated per mass of oxygen utilised by the thermophilic bacteria, is close to unity: The observed respiration quotient for the period was Y_{CO2}=0.70 mol $(CO_2)_{gen}/mol(O_2)_{ut}$ which in terms of mass, is equivalent to 0.975kg$(CO_2)_{gen}$/kg$(O_2)_{ut}$. Accepting that the dry air mass flow rate remains constant through the system, Eq 3.97 in terms of gas stream humidities is as follows:

$$M(H_2O)_{vap} = M(AIR)_{in}(U_{out} - U_{in}) \quad \ldots \text{kg/h} \quad (3.100)$$

3.5.5 Quantifying the Humidity in Terms of Water Vapour Partial Pressure

For the purposes of this particular development, the gas stream is considered to be an ideal gas consisting of two component gases; water vapour and dry air. Application of the ideal gas equation to the effluent and influent gas stream yields:

$$\frac{Q(GAS)_{out}}{R_0.T(GAS)_{out}} = \frac{M(H_2O)_{out}}{m_w(H_2O).P(H_2O)_{out}} = \frac{M(AIR)_{out}}{m_w(AIR)_{out}.P(AIR)_{out}} \quad \ldots (3.101)$$

$$\frac{Q(GAS)_{in}}{R_0.T(GAS)_{in}} = \frac{M(H_2O)_{in}}{m_w(H_2O).P(H_2O)_{in}} = \frac{M(AIR)_{in}}{m_w(AIR)_{in}.P(AIR)_{in}} \quad \ldots (3.102)$$

where:

R_0 = Ideal Gas Constant (mmHg.m^3/kmol.K)
$Q(GAS)_{in,out}$ = Volumetric flow rate of the gas stream (dry air + water vapour) (m^3/h)
$T(GAS)_{in,out}$ = Absolute temperature of the gas stream (K)
$P(H_2O)_{in,out}$ = Partial pressure exerted by the water vapour in the gas stream (mmHg)
$P(AIR)_{in}$ = Partial pressure exerted by the dry air in the gas stream (mmHg)
$m_w(H_2O)$ = Molecular mass of water (kg/kmol)
$m_w(AIR)_{in,out}$ = Molecular mass of dry air in the gas stream (kg/kmol)
$()_{in,out}$ = Subscripts denoting influent or effluent gas streams respectively

Consistent with the previous assumption that the dry air mass flow rate remains constant through the system, it is accepted that the molecular mass of the dry air also remains unchanged through the system and is taken to be that of dry atmospheric air ($m_w(AIR)$ = 29.0 kg/kmol) (extracted from Mayhew and Rogers 1977). Substituting for both the molecular mass of dry air and water into the Ideal Gas Equation for the effluent gas Eq 3.101 and influent gas Eq 3.102 streams yields:

$$\frac{M(H_2O)_{out}}{18P(H_2O)_{out}} = \frac{M(AIR)_{out}}{29P(AIR)_{out}} \quad \ldots (3.103)$$

$$\frac{M(H_2O)_{in}}{18P(H_2O)_{in}} = \frac{M(AIR)_{in}}{29P(AIR)_{in}} \quad \ldots (3.104)$$

In terms of influent and effluent gas stream humidity the above equations become:

$$U_{out} = \frac{18}{29} \cdot \frac{P(H_2O)_{out}}{P(AIR)_{out}} \quad \ldots (3.105)$$

$$U_{in} = \frac{18}{29} \cdot \frac{P(H_2O)_{in}}{P(AIR)_{in}} \quad \ldots (3.106)$$

For an ideal gas the total pressure exerted by a gaseous mixture is equal to the sum of the individual component partial pressures (Dalton's Law). For the aerobic reactor gas stream, this can be expressed mathematically by:

$$P(GAS)_{in} = P(AIR)_{in} + P(H_2O)_{in} \quad \ldots mmhg \quad (3.107)$$

$$P(GAS)_{out} = P(AIR)_{out} + P(H_2O)_{out} \quad \ldots mmHg \quad (3.108)$$

The total pressure on the system as the gas stream is vented from the reactor is assumed constant at atmospheric pressure i.e. 760 mmHg (1 atm. or 101.3 kPa). The total pressure on the influent gas stream on entry to the aerobic reactor is assumed to be equal to atmospheric pressure plus the pressure exerted by the head of water/sludge in the aerobic reactor. The pressure exerted by a head of water 10.3m in height is 760mmHg. The head of water/sludge in the aerobic reactor at the point of influent gas release is 6.5 m. It follows therefore that the pressure at the point of entry into the reactor is equal to 760 x (1 + 6.5/10.3). The total pressure on the effluent and influent gas streams are therefore as follows:

$$P(GAS)_{out} = 760 \quad \ldots mmHg \quad (3.109)$$

$$P(GAS)_{in} = 1240 \quad \ldots mmHg \quad (3.110)$$

Substituting the above values for total gas stream pressure (influent and effluent) into Eq 3.107 and Eq 3.108, rearranged in terms of dry air partial pressure yields:

$$P(AIR)_{out} = 760 - P(H_2O)_{out} \quad \ldots mmHg \quad (3.111)$$

$$P(AIR)_{in} = 1240 - P(H_2O)_{in} \quad \ldots mmHg \quad (3.112)$$

Substituting the above expressions for dry air partial pressure into Eq 3.105 and Eq 3.106 yields:

$$U_{out} = \frac{18}{29} \cdot \frac{P(H_2O)_{out}}{760 - P(H_2O)_{out}} \quad \ldots (3.113)$$

$$U_{in} = \frac{18}{29} \cdot \frac{P(H_2O)_{in}}{1240 - P(H_2O)_{in}} \quad \ldots (3.114)$$

Equations Eq 3.113 and Eq 3.114 express the humidity of the gas stream in terms of the partial pressure exerted by the water vapour in the gas stream. The partial pressure is a function of the gas stream temperature, and therefore its dependence on the temperature needs to be quantified. It should be noted that at this stage, the Equations Eq 3.113 and Eq 3.114 for humidity are valid for all degrees of water vapour saturation in the gas stream and are not specific to the case where the gas stream is fully saturated.

3.5.6 Water Vapour Partial Pressure as a Function of Temperature

Provided there is sufficient contact time, as the gas stream passes through the sludge liquid, water vapour is transferred from the sludge liquid to the gas phase until the partial pressure of the water vapour in the gas stream equals the vapour pressure of the sludge liquid. When the partial pressure of the water vapour equals the vapour pressure of the sludge liquid, the gas stream is saturated. Any subsequent fluctuation in the vapour pressure (through a change in the sludge liquid temperature) will bring about a corresponding change in the water vapour partial pressure to maintain equilibrium; consequently there will be a change in gas stream humidity (Eq 3.113 and Eq 3.114).

The dominant influence on the vapour pressure, and consequently the partial pressure of water vapour in the gas stream, is the temperature of the sludge liquid. The vapour pressure of water is known to increase rapidly with an increase in liquid temperature; a variation which is expressed quantitatively by the Clapeyron Equation (Eq 3.115). If it can be accepted that both the influent and effluent gas streams are saturated, then the Clapeyron equation can be employed for the determination of the water vapour partial pressure in terms of gas temperature.

For the Athlone aerobic reactor (air oxygenated), it is reasonable to assume that both the influent and effluent gas streams are fully saturated at the influent and effluent gas stream temperatures $T(GAS)_{in}$ and $T(GAS)_{out}$ respectively. The influent gas stream can be considered saturated after passage through the liquid ring compressor because

significant quantities of water were continually being drained from the influent gas stream pipeline. The effluent gas stream can be considered saturated after passage upwards through the aerobic reactor sludge because with the gas cooling that took place in the reactor head space in the absence of a significant foam layer significant quantities of condensate were observed; this would not have been evident if the effluent gas stream were not saturated. Since it was reasonable to accept that the vent gas was saturated with water vapour, it could also be assumed that there was sufficient contact time between the sludge liquid and the gas stream, for equilibrium to be reached in terms of the transfer of sensible heat between the two phases. Accordingly the temperature of the gas stream as is it escaped the surface of the sludge liquid was accepted to be at the sludge liquid temperature. It should be noted that, under non-foaming conditions the gas stream undergoes some degree of cooling in the reactor head space and exits the reactor at a lower temperature than the sludge in the reactor (-3°C). A large proportion of the vapour heat given up by the gas stream is lost from the system through the walls of the reactor head space (see Section 3.7). The humidity of the gas stream as it exits the reactor (and therefore the vapour heat loss rate) is therefore defined in terms of the temperature of the gas as it exits the reactor ($T(AIR)_{out}$), which is a measurable parameter.

For a two phase (liquid-gas) system, in equilibrium in terms of mass and heat transfer, the Clapeyron Equation is as follows:

$$\frac{dP^{sat}}{dT} = \frac{ML_v}{T\,(V_G - V_L)} \quad \text{...MJ/kg} \quad (3.115)$$

where:

dP^{sat}/dT = Rate of change of saturation pressure with temperature (MJ/m^3.K)
M = Mass of substance vaporised (kg)
L_v = Latent heat of vaporisation (MJ/kg)
$(V_G - V_L)$ = Change in volume accompanying the phase change (m^3)
T = Absolute temperature at which the phase change occurs (K)

The use of the Clapeyron equation in this form is, however, greatly restricted because it presupposes a knowledge of L_v, V_G and V_L and their variation with temperature. For vaporisation processes at low pressures (as is the case with the process in the aerobic reactor) one may introduce reasonable approximations into the Clapeyron equation by assuming that (1) the vapour phase is an ideal gas and (2) the volume of liquid vaporised is considered small with respect to the volume of vapour generated by the volume lost (for water V_L <0.1% of V_G). The above assumptions can be expressed by:

$$V_G - V_L = V_G = \frac{MR_0T}{m_w P^{sat}} \quad \text{...m}^3 \quad (3.116)$$

Upon substitution of the above, the Clapeyron equation reduces to the familiar Claussius-Clapeyron equation:

$$\frac{dP^{sat}}{dT} = \frac{L_v . m_w . P^{sat}}{R_0 . T^2} \quad \text{...MJ/m}^3\text{.K} \quad (3.117)$$

The latent heat of vaporisation decreases with increasing temperature, making an exact solution to the above equation extremely complex. However, for a system where the temperature does not vary over wide limits, it may be assumed that the latent heat of vaporisation remains constant; an assumption made earlier. The variation in the temperature range encountered in both the influent and effluent gas streams (15-25°C and 45-65°C respectively) for the aerobic reactor is sufficiently small to allow this assumption. Constant values for the latent heat of vaporisation, of the effluent and influent gas streams respectively, can therefore be accepted with negligible loss of accuracy. Rearranging the Claussius-Clapeyron equation in the following form:

$$\frac{dP^{sat}}{P^{sat}} = \frac{m_w L_v}{R_0} \frac{dT}{T^2} \quad \text{...} \quad (3.118)$$

and integrating between the limits P_0^{sat}, T_0 and P^{sat}, T yields:

$$\ln P^{sat} - \ln P_0^{sat} = \frac{m_w L_v}{R_0} \left(\frac{1}{T_0} - \frac{1}{T} \right) \quad \text{...} \quad (3.119)$$

Equation Eq 3.119 permits calculation of the saturation vapour pressure P^{sat} of a substance at a temperature T if the vapour pressure P_0^{sat} at another temperature T_0 is known together with the latent heat of vaporisation L_v. Treating the influent and effluent air cases separately, the following data was extracted from the Steam Tables by Mayhew and Rogers (1977):

Table 3.4 Latent Heat and Water Vapour Partial Pressure Data at 20°C and 50°C

Parameter	Influent Air	Effluent Air
Reference Temperature (T_0) Latent Heat (L_v) Partial Pressure (P_0^{sat})	(273 + 20°C) = 293K 2454 kJ/kg(H_20) 17.8 mmHg	(273 + 50°C) = 323K 2382 kJ/kg(H_2O) 93.8 mmHg
R_0 = 8.314 kJ/kmol/K	In(x) = 2.303 $\log_{10}$(x)	

Substituting the above values into Eq 3.119 yields:

$$\log_{10}\left(P(H_2O)_{out}^{sat}\right) = \frac{18 \times 2382}{2.303 \times 8.314}\left(\frac{1}{323} - \frac{1}{273 + T(AIR)_{out}}\right) + \quad \ldots (3.120)$$

$$\log_{10}\left(P(H_2O)_{in}^{sat}\right) = \frac{18 \times 2454}{2.303 \times 8.314}\left(\frac{1}{293} - \frac{1}{273 + T(AIR)_{in}}\right) + 1 \quad \ldots (3.121)$$

where:

$P(H_2O)^{sat}_{in,out}$ = Water vapour saturation partial pressure in the influent and effluent gas stream respectively (mmHg)

$T(AIR)_{out}$ = Temperature of the influent and effluent gas stream respectively (°C)

Simplification of Eq 3.120 and Eq 3.121, gives the saturation partial pressure of water vapour in terms of the gas stream temperature for both effluent and influent gas streams respectively, ie:

$$P(H_2O)_{out}^{sat} = a\log_{10}\left(8.903 - \frac{2239}{273 + T(AIR)_{out}}\right) \quad \ldots\text{mmHg} \quad (3.122)$$

$$P(H_2O)_{in}^{sat} = a\log_{10}\left(9.124 - \frac{2307}{273 + T(AIR)_{in}}\right) \quad \ldots\text{mmHg} \quad (3.123)$$

Substitution of the above equations for effluent (Eq 3.122) and influent (Eq 3.123) water vapour partial pressure into the humidity equations Eq 3.113 and Eq 3.114 respectively, yields the following:

$$U_{out}^{sat} = \frac{18}{29} \cdot \frac{alog_{10}\left(8.903 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.903 - \frac{2239}{273 + T(AIR)_{out}}\right)} \quad \ldots (3.124)$$

$$U_{in}^{sat} = \frac{18}{29} \cdot \frac{alog_{10}\left(9.124 - \frac{2307}{273 + T(AIR)_{in}}\right)}{1240 - alog_{10}\left(9.124 - \frac{2307}{273 + T(AIR)_{in}}\right)} \quad \ldots (3.125)$$

3.5.7 Calculation of the Water Vaporisation Rate

The water vaporisation rate (when oxygenation is with air only) is calculated from the product of the mass flow rate of dry air through the reactor (assumed unchanged) and the difference in the humidity of the influent and the effluent gas streams (Eq 3.100).

The humidity of the influent and effluent gas streams are both expressed as a function of gas stream temperature, given by Eq 3.123 and Eq 3.124 respectively. As the dry air volumetric flow rate $Q(AIR)_{in}$ was the parameter used to measure the quantity of dry air into the aerobic reactor, it is necessary to link the mass flow rate to the volumetric flow rate. This is done via the density:

$$M(AIR)_{in} = 1.205 \times Q(AIR)_{in} \qquad (see\ Eq.\ 2.2) \qquad \ldots kg/h \quad (3.126)$$

where:

1.205 = Dry air density at 20°C and 760mmHg (kg/m^3)

$Q(AIR)_{in}$ = Volumetric flow rate of dry air in the influent gas stream (m^3(AIR)/h)

Substitution of Eq 3.124, Eq 3.125 and Eq 3.126 into Eq 3.100 for the mass rate of water vaporisation yields:

$$M(H_2O)_{vap} = 1.205\,Q(AIR)_{in}\left(\frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{0.62 \times alog_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)}{1240 - alog_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)}\right) \quad \ldots kg/h \quad (3.127)$$

A graphical representation of Equation 3.127 for the rate of water vaporisation at saturation versus the effluent gas temperature $T(AIR)_{out}$ for different influent air flow rates $Q(AIR)_{in}$ is given in Figure 3.7 below.

From Figure 3.7, it can be seen that at an air flow rate $Q(AIR)_{in}$ of 760m^3(STP)/h, and reactor sludge temperature of 55°C, the water vaporisation rate is about 100kg(H_2O)/h

or equivalently about 100l/h. In a day this amounts to 2.4m^3 which from the aerobic reactor volume of 184m^3 constitutes a loss of (100×2.4/184) 1.3%. At influent flows of 45 and 30m^3/d (4 and 6 days retention time respectively), 2.4m^3/d constitutes a loss of 5.3 and 8.0% of the influent flow rate respectively (see Section 2.4.4).

3.5.8 Calculation for the Vapour Heat Loss Rate: Oxygenation with Air Only

The vapour heat loss rate H_v in the effluent gas stream, was defined (Eq 3.96) as the product of the latent heat of vaporisation (Eq. 3.95) and the water vaporisation rate (Eq 3.100 and Eq 3.127). From the derivation, it follows that the water vapour heat loss rate when oxygenation is with air only is given by:

$$H_v = L_v^{50°C}\ \rho(AIR)\,Q(AIR)_{in}\,(U_{out} - U_{in}) \quad \text{...MJ/h} \quad (3.128)$$

The expression for the water vapour heat loss rate H_v in terms of the measurable parameters (1) volumetric air flow rate $Q(AIR)_{in}$ and (2) influent and effluent gas stream temperatures $T(AIR)_{in}$ and $T(AIR)_{out}$, is as follows:

$$H_v = 2.38 \times 1.205\,Q(AIR)_{in}\left(\frac{0.62a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{0.62a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}{1240-a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}\right)$$

Equation 3.129 The Water Vapour Heat Loss Rate in the Effluent Gas H_v (MJ/h) when Oxygenation is with Air Only (Phase I)

Note that the expression is only valid when oxygenation is with air alone and both the influent and effluent gas streams are fully saturated with water vapour. Water vapour saturation of the influent gas stream was indirectly verified for the Athlone aerobic reactor by observing large quantities of condensate being discharged from the delivery end of the liquid ring compressor. Saturation of the effluent gas stream was also indirectly verified by (1) observing large quantities of condensate on the inner walls of the reactor head space (see Section 3.4.5) and (2) through the finding that the estimated specific heat yield Y_h was very close to that obtained by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor (see Section 4.3.8).

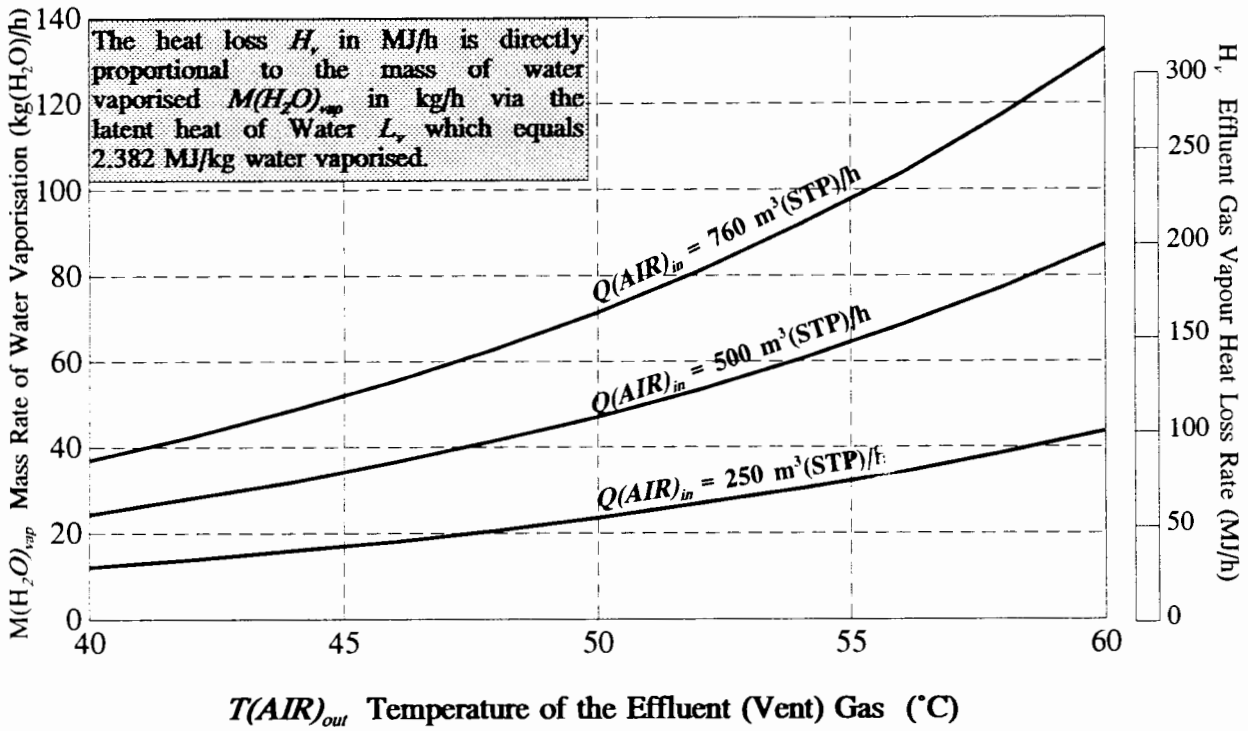

Figure 3.7 **Mass Rate of Water Vaporisation $M(H_2O)_{vap}$ and the Water Vapour Heat Loss Rate H_v in the Effluent Gas versus Effluent (Vent) Gas Temperature $T(AIR)_{out}$ for different Influent Air Flow Rates $Q(AIR)_{in}$, from Equations 3.127 and 3.129 respectively. The Influent Air Flow Temperature is taken to be 20°C. Oxygenation is with Air Alone.**

3.5.9 Calculation of the Vapour Heat Loss Rate During Pure Oxygen Supplementation

The injection of pure oxygen into the sludge recirculation line to supplement the aeration system (performed during phase II) increases the flowrate of gas through the system. The pure oxygen influent mass flow rate $M(O_2)_{in}^{O2}$ was measured directly with a rotameter flow meter. The influent mass flow rate of air $M(AIR)_{in}$ was given above by Eq 3.126 in terms of the dry air density $\rho(AIR)$ and the influent volumetric flow rate of air $Q(AIR)_{in}$ (which was measurable) viz:

$$M(AIR)_{in} = \rho(AIR).Q(AIR)_{in} = 1.205 \times Q(AIR)_{in} \quad \text{...kg/h} \quad (3.126)$$

The combined influent gas mass flow rate $M(GAS)_{in}$ is therefore given by:

$$M(GAS)_{in} = M(AIR)_{in} + M(O_2)_{in}^{O2} = 1.205 \times Q(AIR)_{in} + M(O_2)_{in}^{O2} \quad \text{...kg/h} \quad (3.131)$$

Note that, as in the development of H_v above (for oxygenation with air alone), the effluent gas mass flow rate $M(GAS)_{out}$ is accepted to be equal to the influent gas mass flow rate $M(GAS)_{in}$ for oxygenation with both air and pure oxygen. This assumption is

valid as the observed respiration quotient for phase I was Y_{CO2}=0.70mol$(CO_2)_{gen}$/mol$(O_2)_{ut}$ which in terms of mass, is equivalent to 0.975kg$(CO_2)_{gen}$/kg$(O_2)_{ut}$.

Accordingly the equation (Eq 3.128) for the vapour heat loss rate H_v^{AIR+O2}, when oxygenation is with air and pure oxygen, takes the form:

$$H_v^{AIR+O2} = L_v^{50°C}.\left(\rho(AIR)\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left(U_{out} - U_{in}\right) \quad \text{...MJ/h} \quad (3.132)$$

The expression for the water vapour heat loss rate H_v^{AIR+O2} in terms of the measurable parameters (1) the volumetric air flow rate $Q(AIR)_{in}$ and mass pure oxygen flow rate $M(O_2)_{in}^{O2}$ and (2) the influent and effluent gas stream temperatures $T(AIR)_{in}$ and $T(AIR)_{out}$, is as follows:

$$H_v^{AIR+O2} = 2.38\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left(\frac{0.62a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{0.62a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}{1240-a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}\right)$$

Equation 3.133 The Water Vapour Heat Loss Rate in the Effluent Gas H_v^{AIR+O2} (MJ/h) during Pure Oxygen Supplementation (Phase II)

Note that, as with Eq 3.129 above, the expression is only valid when both the influent and effluent gas streams are fully saturated with water vapour.

The effect of injecting pure oxygen into the reactor has a relatively minimal effect on the vapour heat loss rate in comparison to the amount of additional biological heat with which it is able to generate. For example, at an influent air flow rate of 760m^3(STP)/h, corresponding to 915kg(AIR)/h or 211kg(O_2)/h, which at an ***OTE*** of 12.8% yields an oxygen transfer rate or oxygen utilisation rate of 27.0kg(O_2)/h. An additional influent pure oxygen mass flow rate of 91.5 kg(O_2)/h would increase the vapour heat loss rate by only 10%. However, the biological heating rate would increase from approximately **350MJ/h** (13MJ/kg(O_2) ×27kg(O_2)/h) (refer Section 4.3.7 below; steady state periods 3,7 and 8) to over **1300MJ/h** ((80%×91.5+27)kg(O_2)/h×13MJ/kg(O_2)) at a pure oxygen transfer efficiency of 80%. An increase of **270%**.

3.6 THE SENSIBLE HEAT LOSS IN THE EFFLUENT GAS

3.6.1 Introduction

The heat which is transferred from the hot sludge to the cool gas stream as the gas passes upwards through the aerobic reactor is ultimately lost from the system when the gas is vented. This energy loss is termed the sensible heat loss in the effluent gas H_g.

In order to quantify the sensible heat loss in the effluent gas H_g consideration needs to be given to the changes in heat content of the individual components associated with the gas stream as it passes through the system. For any given substance the change in its heat content is dependent upon (1) its heat capacity, (2) its mass, and (3) the change in temperature.

The derivation of an expression for H_g looks at (1) the application of heat capacities for the individual components in the gas stream (2) the changes in mass which take place in the components of the gas stream and (3) the temperature changes which are effected. The Section concludes with the derivation of an expression which incorporates the increase in gas sensible heat loss due to the increase in vent gas flow rate during pure oxygen supplementation.

3.6.2 Defining Heat Capacity

In general terms heat capacity is defined as the amount of heat required to increase the temperature of a given substance by one degree. Mathematically this can be expressed by:

$$C = \frac{1}{m} \cdot \frac{\delta q}{\delta T} \quad \text{...MJ/kg.K} \quad (3.134)$$

where:

C = The heat capacity of the substance heated (Mk/kg.K)

m = The mass of the substance heated (kg)

δq = The quantity of heat transferred to the substance (MJ)

δT = The temperature rise effected by the transfer of heat δq (K)

As it stands this definition is not complete, because the manner in which the heat δq is transferred needs to be specified, ie the pressure, volume and temperature constraints on the system need to be identified.

For a system at constant pressure, the heat transferred to the substance is equal to the change in the heat content of that substance ($\delta q = dH$). The change in heat content in terms of the heat capacity of the substance is therefore given by:

$$dH_p = mC_p dT \quad \text{...MJ} \quad (3.135)$$

where:

dH_p = The change in heat content at constant pressure (MJ)

C_p = The Specific Heat Capacity at constant pressure (MJ/kg.K)

However, for the aerobic reactor at Athlone, the gas (air) stream pressure does not remain constant; The influent gas (air) stream pressure $P(GAS)_{in}$ is higher than the effluent gas stream pressure[1] $P(GAS)_{out}$, ie:

$$P(GAS)_{in} = 1.63 \quad (see\ Eq\ 3.56) \quad \text{...atm} \quad (3.136)$$

$$P(GAS)_{out} = 1.00 \quad (see\ Eq\ 3.55) \quad \text{...atm} \quad (3.137)$$

The effect of this change in pressure on the heat content of the gas stream can be evaluated by considering the Joule-Thomson effect and applying it to the conditions prevalent at Athlone.

3.6.3 Considering the Joule-Thompson Effect

The change in temperature of a gas as it passes from a region of higher pressure into a region of lower pressure, assuming the whole system to be lagged so that the expansion is adiabatic (without heat loss), is described by the Joule-Thomson Equation:

$$\mu_{JT} = \left(\frac{dT}{dP}\right)_H \quad \text{...°C/atm} \quad (3.138)$$

where:

μ_{JT} = The Joule Thompson Coefficient (°C/atm)

dT = The change in Temperature (°C)

dP = The change in Pressure (atm)

$[\,]_H$ = Signifies constant enthalpy (no heat loss)

---

[1] The term $P(GAS)_{in,out}$ is used (as opposed to $P(AIR)_{in,out}$) because it represents the pressure exerted by both the dry air and the water vapour components in the gas stream (refer to Equations 3.54 and 3.55).

To estimate the change in heat content of the gas stream as a result of the decrease in pressure through the system, the following values for air were applied to the Joules-Thomson equation Eq 3.138 arranged in terms of dT: (1) at a temperature of 50°C and system pressure of 1 atm, μ_{JT} = 0.189 (Perry and Chilton, 1973), and (2) the gas stream pressure dropped by approximately 0.63 atm in passing through the aerobic reactor (Eq 3.136 and Eq 3.137). Hence the changes in temperature dT as result of the reduction in pressure is:

$$dT = \mu_{JT} dP = 0.189 \times 0.63 = 0.120 \quad ...°C \quad (3.139)$$

This temperature change of about 0.1°C is negligible in comparison with the temperature increase in the gas stream effected by the transfer of heat from contact with the hot sludge (i.e. from 20°C up to 50° to 60°C), with the result that the change in heat content of the gas stream as a result of the change in gas pressure can therefore be ignored and Eq 3.135 can be directly applied. To calculate the change in heat content for a given substance, as a result of a finite change in temperature between T_1 and T_2, Eq 3.135 is integrated between the limits T_1 and T_2:

$$\Delta H = \int_{T_1}^{T_2} m C_p dT \quad ...MJ \quad (3.140)$$

3.6.4 Effect of Temperature on the Heat Capacity

The heat capacity C_p of a gas is dependent upon its temperature, making the solution to Eq 3.138 extremely complex if the effect is severe. However, if the heat capacities of the various components associated with the gas stream can be shown to be relatively constant over the temperature range encountered in the aerobic reactor (ie independent of temperature) then the above integral of Eq 3.140 can be readily evaluated. For the components in the gas stream the specific heats at 20°C and 50°C were obtained from Mayhew and Rogers (1977) and are given in Table 3.5

Table 3.5 Heat Capacities for the Components in the Gas Stream

Gas	Heat Capacity (MJ/kg.°C)	
	C_p at 20°C	C_p at 50°C
Oxygen	0.918×10^{-3}	0.923×10^{-3}
Nitrogen	1.040×10^{-3}	1.040×10^{-3}
Carbon Dioxide	0.837×10^{-3}	0.869×10^{-3}
Water Vapour	1.863×10^{-3}	1.870×10^{-3}

From the above, it is quite reasonable to accept constant values for the specific heats for each of the components in the gas stream. Consequently Eq 3.140 can be simplified to:

$$\Delta H = m.C_p.(T_2 - T_1) \quad \text{...MJ} \quad (3.141)$$

3.6.5 Consideration of Mass Transfer

Solution for the sensible heat loss is complicated by the fact that the composition of the gas stream does not remain constant. Mass transfer takes place between the sludge liquid phase and the gas phase from the effects of (1) aeration, (2) biological respiration and (3) water vaporisation. The principle effects of this interaction are as follows:

(1) A certain percentage of the oxygen in the influent gas stream is transferred into solution as the gas stream passes through the sludge (defined by the oxygen transfer efficiency *OTE).*

(2) The oxygen transferred into solution is utilised by the thermophilic bacteria and through biological respiration, carbon dioxide is released into the gas stream. The respiration quotient Y_{CO2}, which describes the number of moles of carbon dioxide generated per mole of oxygen utilised, was measured at 0.70 kmol/kmol during the evaluation period.

(3) The quantity of water vapour in the gas stream increases as the gas stream passes upwards through the aerobic reactor, and is saturated when the gas stream is vented from the reactor. Consequently the quantity of effluent sludge is reduced as a result of the loss of water vapour to the gas stream (by about 5 to 8% depending on temperature and retention time).

3.6.6 Simplifying the Calculation for the Gas Sensible Heat Loss Rate

To simplify the eventual solution for the sensible heat loss in the effluent gas, six assumptions and approximations are made. Stepwise the six are as follows:

(1) The aerobic reactor gas stream is considered to consist of two component gases; dry air (of variable composition) and water vapour. The total sensible heat loss rate from the aerobic reactor in the effluent gas stream can therefore be expressed by:

$$H_g = H_g(AIR) + H_g(H_2O) \quad \text{...MJ/h} \quad (3.142)$$

where:

H_g = The total sensible heat loss rate with the effluent gas stream (MJ/h)
$H_g(AIR)$ = The sensible heat loss rate with the dry air in the effluent gas (MJ/h)
$H_g(H_2O)$ = The sensible heat loss rate with the water vapour in the effluent gas (MJ/h)

(2) The mass flow rate of dry air through the system is assumed to remain constant, an assumption already made and explained earlier (see Section 3.5.4). The mass flow rate of dry air can be expressed in terms of the volumetric flow rate via the density of air (Eq 3.36). The quantity of air supplied to the aerobic reactor was monitored by the volumetric dry air flow rate $Q(AIR)_{in}$ corrected to 20°C and 760mmHg(STP), at which temperature and pressure the dry air density of air is computed to be ρAIR = 1.205 kg/m^3. It follows therefore that:

$$M(AIR)_{out} = M(AIR)_{in} = 1.205\ Q(AIR)_{in} \quad \text{...kg/h} \quad (3.143)$$

where:

$M(AIR)_{in,out}$ = Mass flow rate of dry influent and effluent air respectively (kg/h)
1.205 = Density of dry air at STP (kg/m^3)
$Q(AIR)_{in}$ = Volumetric flow rate of dry influent air corrected to STP (m^3)

(3) The heat capacity of the dry air remains constant as the gas stream passes through the system, in-spite of the biological respiration which takes place, and can be accepted to be equal to that of atmospheric air at 20°C. From Mayhew and Rogers (1977) $C_p^{20°}(AIR)$ = 1.00 kJ/kg.°C. This assumption is based on (i) the relative similarity between the specific heats of carbon dioxide and oxygen (see Table 3.5) and (ii) the minimal change in C_p values for the individual component gases over the temperature range encountered in the aerobic reactor (see Table 3.5). In equation form this assumption can be expressed by:

$$C_p(AIR)_{out} = C_p(AIR)_{in} = C_p^{20°C}(AIR) = 1.00 \times 10^{-3} \quad \text{...MJ/kg.°C} \quad (3.144)$$

where:

$C_p^{20°C}(AIR)$ = Heat capacity of dry atmospheric air at 20°C (MJ/kg.°C)
$C_p(AIR)_{in,out}$ = Heat capacity of dry influent and effluent air (MJ/kg.°C)

(4) All the water vapour in the effluent gas stream was obtained through the vaporisation of water from the liquid sludge; ie the contribution from the humidity (water vapour) in the influent gas stream is negligible. The humidity of the influent gas stream at 20°C is $U_{in}^{20°C}$ = 0.009 and that for the effluent gas stream at 50°C is $U_{out}^{50°C}$ = 0.087; these estimates were obtained by application of Eq 3.124 and Eq 3.125 assuming gas stream saturation and shows the difference to be about 90%. i.e. 10% of the water vapour in the effluent gas stream was present in the influent gas stream, with the remaining 90% being generated through vaporisation in the aerobic reactor. Even though this may seem like a large error in that the water vapour mass flow rate is overestimated by 10%, it should be noted that this overestimation has a negligible effect on the heat balance i.e. <1%. This is because firstly the water vapour sensible heat loss is only about 10% of the total gas sensible heat loss and secondly the total gas sensible heat loss is only 5 to 10% of the total heat losses. Therefore ignoring the influent humidity, the mass flow rate of water vapour in the effluent gas stream in terms of effluent humidity is given by:

$$M(H_2O)_{out} = M(H_2O)_{vap} = M(AIR)_{in}.U_{out} \quad \text{...kg/h} \quad (3.145)$$

where:
$M(H_2O)_{out}$ = Mass flow rate of water vapour in the effluent air stream (kg/h)
$M(H_2O)_{vap}$ = Mass flow rate of water vaporised from the sludge stream (kg/h)
$M(AIR)_{in}$ = Mass flow rate of dry air through the aerobic reactor (kg/h)
U_{out} = Humidity of the effluent gas stream

(5) The effluent gas stream is assumed to be saturated at the vent gas temperature after passage through the liquid sludge. The humidity of the effluent gas is (from Eq 3.128):

$$U_{out} = \frac{0.62 \times a\log_{10}\left(8.903 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.903 - \frac{2239}{273 + T(AIR)_{out}}\right)} \quad \ldots \quad (3.146)$$

(6) The heat capacity of the water vapour in the effluent gas stream is assumed to remain constant over the temperature range encountered (Table 3.5). From Mayhew and Rogers (1977):

$$C_p^{50°C}(H_2O)_{vap} = 1.87 \times 10^{-3} \quad \text{...MJ/kg.°C} \quad (3.147)$$

where:
$C_p^{50°C}(H_2O)_{vap}$ = Heat capacity of water vapour at 50°C (MJ/kg.°C)

The heat capacity of water vapour at 50°C was selected as this was the average temperature during phase I, when the fraction of the feed sludge volume vaporised inside the reactor was at its highest (5 to 8%). There is however negligible change in the heat capacity at 65°C ($C_p^{65°C}(H_2O)_{vap}$=1.88kJ/kg.°C).

3.6.7 Calculation of the Gas Sensible Heat Loss Rate

The sensible heat loss with the dry air and water vapour fractions in the effluent gas stream are given respectively by:

$$H_g(AIR) = M(AIR)_{in} . C_p^{20°C}(AIR) (T(AIR)_{out} - T(AIR)_{in}) \quad \text{...MJ/h} \quad (3.148)$$

$$H_g(H_2O) = M(AIR)_{in} . U_{out} . C_p^{50°C}(H_2O)_{vap} . T(AIR)_{out} \quad \text{...MJ/h} \quad (3.149)$$

where:

$T(AIR)_{in}$ = Influent gas stream temperature (°C)

$T(AIR)_{out}$ = Effluent gas stream temperature (°C)

Note that the sensible heat input to the system by the fraction of water in the sludge stream which is subsequently vaporised into the gas stream, is accounted for in the calculation of the sensible heat loss with the sludge (Eq 3.159). Equations Eq 3.148 and Eq 3.149 can now be combined to give the total sensible heat loss in the effluent gas stream:

$$H_g = M(AIR)_{in}\left(C_p^{20°C}(AIR)(T(AIR)_{out} - T(AIR)_{in}) + U_{out} . C_p^{50°C}(H_2O)_{vap} . T(AIR)_{out}\right)$$

...MJ/h (3.150)

The following substitutions are now made into Eq 3.150 above: (1) Eq 3.143 to convert the mass flow rate to the measured parameter volumetric flow rate, (2) Eq 3.144 and Eq 3.148 to apply values to the heat capacities $C_p^{20°C}(AIR)$ and $C_p^{50°C}(H_2O)_{vap}$ respectively, and (3) Eq 3.146 to express humidity in terms of the effluent gas temperature which was measured. The resulting expression for the total sensible heat loss in the effluent gas is given by Eq 3.151 below.

$$H_g = 1.21 Q(AIR)_{in}\left[0.001(T(AIR)_{out} - T(AIR)_{in}) + \frac{0.62 a \log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a \log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} 0.00187 T(AIR)_{out}\right]$$

Equation 3.151 The Effluent Gas Sensible Heat Loss Rate H_g (MJ/h) when Oxygenation is with Air Alone (Phase I)

To give an impression of the sensible heat loss with the saturated effluent (vent) gas stream, Equation 3.151 is shown graphically in Figure 3.8 for an influent air temperature

of 20°C. Comparing Figures 3.7 and 3.8 it can be seen that the sensible heat loss is a relatively minor part of the total heat losses with the effluent gas, i.e. <20%; the water vapour heat loss is the greater part i.e. 80%.

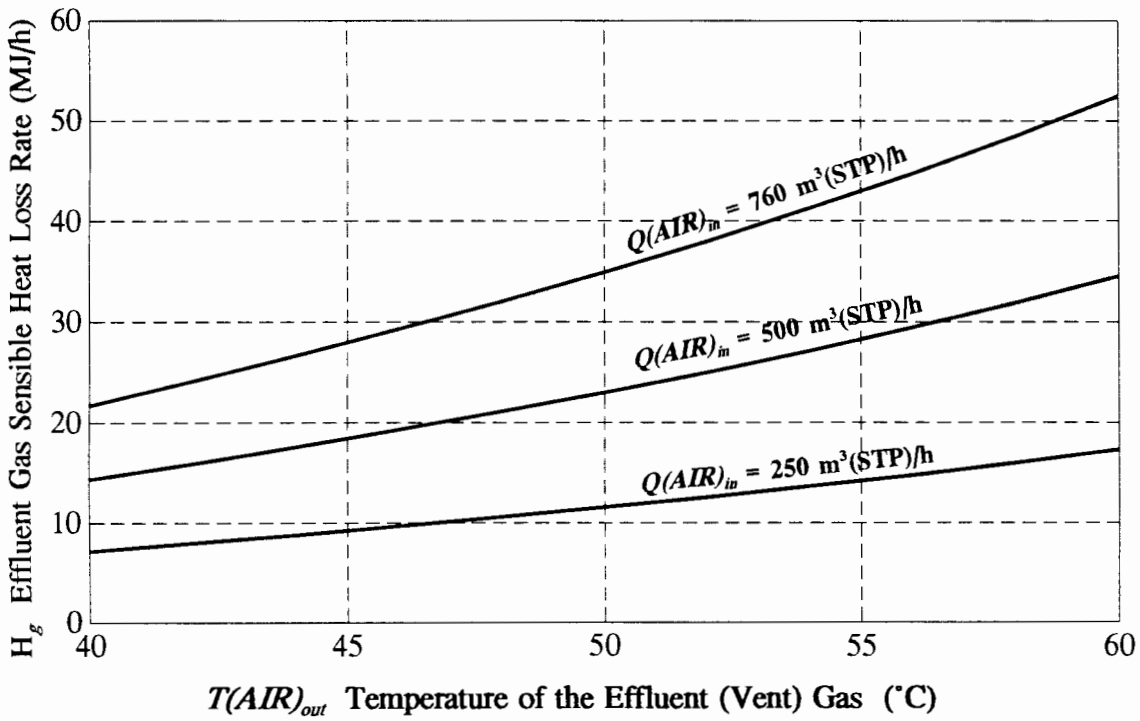

Figure 3.8 The Effluent Gas Sensible Heat Loss Rate (H_g) versus Effluent (Vent) Gas Temperature ($T(AIR)_{out}$) for different Influent Air Flow Rates ($Q(AIR)_{in}$), from Equation 3.151. The Influent Air Flow Temperature is taken to be 20°C. Oxygenation is with Air Alone (Phase I).

3.6.8 Calculation of the Gas Sensible Heat Loss During Pure Oxygen Supplementation

As discussed in Section 3.5.9 above, the injection of pure oxygen into the sludge recirculation line to supplement the aeration system (performed during phase II) increases the flowrate of gas through the system. The combined influent gas mass flow rate $M(GAS)_{in}$ was given above by Eq 3.131 viz:

$$M(GAS)_{in} = M(AIR)_{in} + M(O_2)_{in}^{O2} = 1.205 \times Q(AIR)_{in} + M(O_2)_{in}^{O2} \quad \text{...kg/h} \quad (3.131)$$

Accordingly the equation (Eq 3.150) for the gas sensible heat loss rate (now expressed as H_g^{AIR+O2}) when oxygenation is with air and pure oxygen, takes the form:

$$H_g^{AIR+O2} = \left(M(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left(C_p^{20°C}(AIR)\left(T(AIR)_{out} - T(AIR)_{in}\right) + U_{out}.C_p^{50°C}(H_2O)_{vap}.T(AIR)_{out}\right)$$

...MJ/h (3.152)

The expression for the water vapour heat loss rate H_g^{AIR+O2} in terms of the measurable parameters (1) the volumetric air flow rate $Q(AIR)_{in}$ and mass pure oxygen flow rate $M(O_2)_{in}^{O2}$ and (2) the influent and effluent gas stream temperatures $T(AIR)_{in}$ and $T(AIR)_{out}$, is as follows:

$$H_g^{AIR+O2} = \left(1.21Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left[0.001\left(T(AIR)_{out} - T(AIR)_{in}\right) + \frac{0.62alog_{10}\left(8.90 - \frac{2239}{273+T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273+T(AIR)_{out}}\right)} 0.00187T(AIR)_{out}\right]$$

Equation 3.153 The Effluent Gas Sensible Heat Loss Rate H_g^{AIR+O2} (MJ/h) During Pure Oxygen Supplementation (Phase II)

The above expression is only valid when both the influent and effluent gas streams are fully saturated with water vapour. As in the case of the vapour heat loss H_v^{AIR+O2} (see Section 3.5.9 above for details), the injection of pure oxygen increases the sensible heat loss rate H_g^{AIR+O2} by at most 10%. A level, negligible in comparison to the increase in the biological heating rate.

3.7 THE SENSIBLE HEAT LOSS IN THE EFFLUENT SLUDGE

3.7.1 Introduction

Influent raw sludge is fed into the aerobic reactor at the feed sludge temperature. The sludge is then warmed by the biological and mechanical heat, the former being generated by the thermophilic bacteria and the latter imparted by the mixing pumps. The temperature to which the sludge can be heated is limited by the various heat losses which take place (refer Eq 3.8).

The heat which is lost when the hot sludge is discharged from the reactor is the major source of heat loss from the system, and can account for up to 60 percent of the total heat losses depending on the retention time (the shorter, the greater). This heat loss is termed the sensible heat loss in the effluent sludge H_s and is related to the heat capacity of the sludge, the temperature rise which is effected between the influent and effluent sludge, and the mass flow rate of influent and effluent sludge.

In order to quantify the sensible heat loss rate H_s , consideration therefore needs to be given to (1) assigning an appropriate value for the heat capacity of the sludge, (2) the density of the sludge because feeding is controlled on a volumetric basis, (3) the change in sludge volume as a result of the vaporisation of water within the aerobic reactor, and

(4) the temperature rise which is effected. During pure oxygen supplementation (phase II) there is an increase in the gas flow rate (<10% by mass) which results in a concomitant change in the volume of sludge as a result of vaporisation.

3.7.2 The Heat Capacity of Sewage Sludge

As defined previously by Eq 3.134, the heat capacity of a given substance is the heat required to increase its temperature by one degree. In the case of water, the heat capacity remains relatively constant at 4.18 kJ/kg.°C in the temperature range encountered in the aerobic reactor 15-65°C. In so far as relating the heat capacity of water to that of sewage sludge is concerned, two factors need to be considered; (1) the effect of the volatile (organic) and non-volatile (inorganic) matter contained in the sludge on the heat capacity of the water (the concentration of dry mass sludge solids in the water is less than 6% and therefore that of the water is in excess of 94%, and (2) because sludge feeding is monitored on a volumetric basis, it is more appropriate to express the heat capacity of sludge in volumetric units of MJ/m^3.°C. In order to assign an appropriate value for the heat capacity of the sludge treated in the aerobic reactor during the evaluation period, the following typical composition for the sludge is given in Table 3.6. Quoted also are the different component heat capacity values and densities extracted from Perry and Chilton (1973):

Table 3.6 Evaluation of Sludge Heat Capacity

Component	Mass in 1 m^3 Sludge Kg	Heat Capacity kJ/kg.°C	Density kg/m^3	Heat Capacity MJ/m^3. °C
Water	1000	4.18	1000	4.18
Volatile Solids[1]	40	1.34	690	0.92
Non-Volatile Solids[2]	10	1.32	2100	2.78

[1] cellulose taken as representative of the volatile solids

[2] silica is taken as representative of the non-volatile solids

From the data contained in Table 3.6 the heat capacity of the "typical" sludge is computed to be:

$$C_p^{20°C}(SL) = 4.04 \quad \text{...MJ/m}^3\text{.°C} \quad (3.154)$$

3.7.3 Effect of Vaporisation on the Sensible Heat Loss with the Effluent Sludge

There is a reduction in the volume of aerobic sludge which is discharged from the reactor due to the vaporisation of water into the gas stream. Consequently the sensible heat lost from the system with the effluent sludge stream is reduced in proportion to the mass of water vaporised. If no vaporisation took place within the aerobic reactor then the heat lost from the system through the sensible heat contained in the effluent sludge would be:

$$H_s^* = \frac{Q(SL)_{in}}{24} . C_p^{20°C}(SL) \ (T(SL)_r - T(SL)_{in}) \quad \text{...MJ/h} \quad (3.155)$$

where:

H_s^* = Sensible Heat Loss in the effluent sludge ignoring vaporisation (MJ/h)
$Q(SL)_{in}$ = Volumetric flow rate of sludge into the reactor (m^3/d)
24 = Conversion factor; to convert flow rate from /d to /h
$T(SL)_{in,out}$ = Influent and effluent sludge temperatures (°C)

The mass flow rate of water vaporised from the sludge stream into the gas stream has been derived (see Section 3.5.6) and is given by the approximate relationship:

$$M(H_2O)_{out} = M(H_2O)_{vap} = M(AIR)_{in} . U_{out} \quad \text{...kg/h} \quad (3.156)$$

where:

$M(H_2O)_{out}$ = Mass flow rate of water vapour in the effluent gas (kg/h)
$M(H_2O)_{vap}$ = Mass flow rate of water vaporised from the sludge stream (kg/h)
$M(AIR)_{in}$ = Mass flow rate of dry air through the aerobic reactor (kg/h)
U_{out} = Humidity of the effluent gas stream

Taking the density of water as 1000 kg/m^3, the volume displaced from the sludge by vaporisation into the gas stream is given by:

$$Q(H_2O)_{vap} = \frac{M(H_2O)_{vap}}{1000} = \frac{M(AIR)_{in} . U_{out}}{1000} \quad \text{...m}^3\text{/h} \quad (3.157)$$

where:

$Q(H_2O)_{vap}$ = Rate of volume displacement by vaporisation from the sludge (m^3)

The volumetric flow rate of sludge leaving the reactor is reduced accordingly:

$$\frac{Q(SL)_{out}}{24} = \frac{Q(SL)_{in}}{24} - Q(H_2O)_{vap} = \frac{Q(SL)_{in}}{24} - \frac{M(AIR)_{in}.U_{out}}{1000} \quad ...m^3/h \quad (3.158)$$

Consequently the Equation for the sensible heat loss needs to be modified accordingly:

$$H_s = \frac{C_p^{20°C}(SL)}{24}\left(Q(SL)_{out}.T(SL)_r - Q(SL)_{in}.T(SL)_{in}\right) \quad ...MJ/h \quad (3.159)$$

Substitution for the effluent sludge flow Eq 3.158 yields:

$$H_s = C_p^{20°C}(SL)\left[\left(\frac{Q(SL)_{in}}{24} - \frac{M(AIR)_{in}.U_{out}}{1000}\right)T(SL)_r - \frac{Q(SL)_{in}.T(SL)_{in}}{24}\right] \quad ...MJ/h \quad (3.160)$$

3.7.4 Calculation of the Sludge Sensible Heat Loss Rate

The following substitutions are now made into Eq 3.160 above: (1) Eq 3.145 to convert the mass flow rate to the measured parameter volumetric flow rate, (2) Eq 3.154 for the sludge heat capacity $C_p^{20°C}(SL)$, and (3) Eq 3.124 to express humidity in terms of the effluent gas temperature which was measured. The resulting expression for the total sensible heat loss in the sludge is as follows:

$$H_s = 4.04\left[\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{1.21T(SL)_r Q(AIR)_{in}}{1000}.\frac{0.62a\log_{10}\left(8.90 - \frac{2239}{273+T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.90 - \frac{2239}{273+T(AIR)_{out}}\right)} - \frac{Q(SL)_{in}T(SL)_{in}}{24}\right]$$

Equation 3.161 The Sludge Sensible Heat Loss Rate H_s (MJ/h)

A graphical representation of Equation 3.161 for the rate of sensible heat loss with the effluent sludge versus sludge temperature at different retention times is given in Figure 3.9 below. The reduction in sensible heat loss (as much as 5 to 10%) as a result of water vaporisation taking place in the reactor is indicated on the graph.

As shown above, the water loss through vaporisation was taken into account in calculating the sensible heat loss of the sludge. However, as mentioned in Section 2.4.4 above, this change in flow rate between influent and effluent (5 to 8%) was ignored in the calculation of the retention time and TS,VS and COD reduction values. These calculations were based on the influent flow rate and assumed no change between influent and effluent.

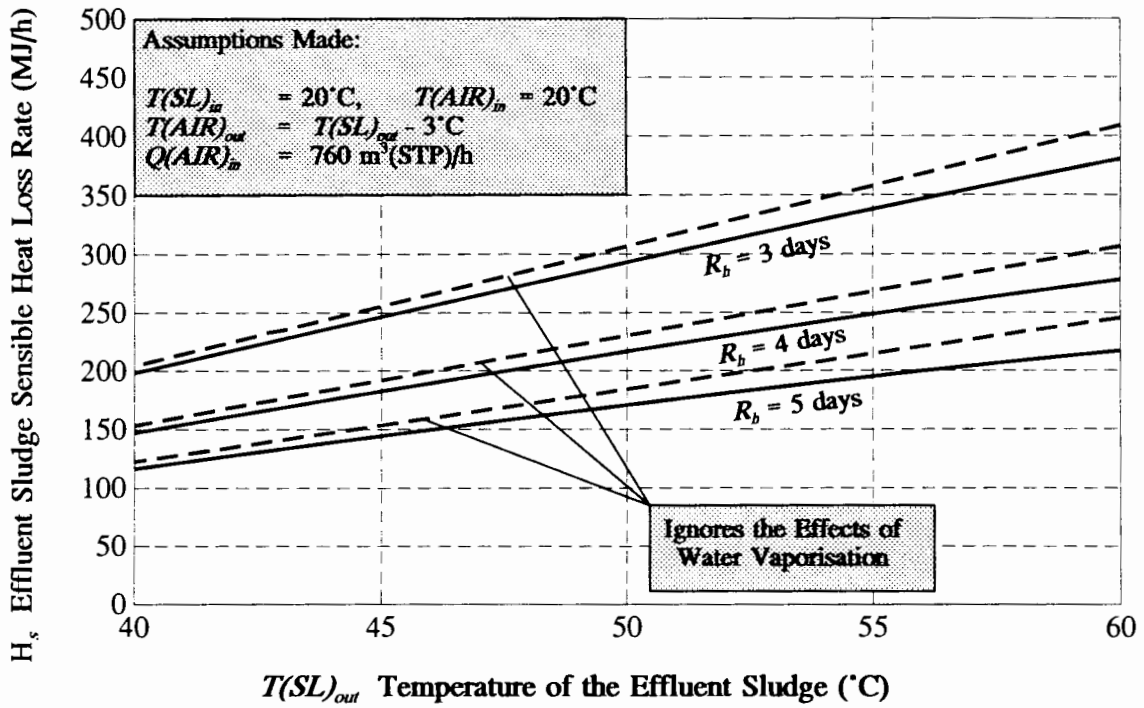

Figure 3.9 **The Effluent Sludge Sensible Heat Loss Rate H_s versus Effluent Sludge Temperature $T(SL)_{out}$ for different Aerobic Reactor Retention Times R_h, from Equation 3.161. The Influent Air Flow and Sludge Temperature are taken to be 20°C.**

3.7.5 Calculation of the Sludge Sensible Heat Loss During Pure Oxygen Supplementation

The injection of pure oxygen into the sludge recirculation line to supplement the aeration system (performed during phase II) increases the flowrate of gas through the system. The combined influent gas mass flow rate $M(GAS)_{in}$ was given above by Eq 3.131 viz:

$$M(GAS)_{in} = M(AIR)_{in} + M(O_2)_{in}^{O2} = 1.205 \times Q(AIR)_{in} + M(O_2)_{in}^{O2} \quad ...kg/h \quad (3.131)$$

Accordingly the expression for the rate of water loss by vaporisation from the sludge $Q(H_2O)_{vap}$ (previously given by 3.167) now takes the form:

$$Q(H_2O)_{vap} = \frac{M(GAS)_{in} . U_{out}}{1000} = \frac{\left(1.205 Q(AIR)_{in} + M(O_2)_{in}^{O2}\right) . U_{out}}{1000} \quad ...m^3/h \quad (3.162)$$

The expression for the sludge sensible heat loss rate H_s^{AIR+O2} during pure oxygen supplementation is therefore as follows:

$$H_s^{AIR+O2} = 4.04\left[\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{T(SL)_r\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right)}{1000} \cdot \frac{0.62 a \log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{Q(SL)_{in} T(SL)_{in}}{24}\right]$$

Equation 3.163 The Sludge Sensible Heat Loss Rate H_s^{AIR+O2} (MJ/h) During Pure Oxygen Supplementation (Phase II)

As discussed in Section 3.7.4 above, the overall reduction in the sensible heat loss through water vaporisation is of the order of 5-10% (Section 3.7.4). The increased gas flow rate as a result of pure oxygen supplementation only increases the rate of water vaporisation by at most 10%. The effect of pure oxygen supplementation on the sludge sensible heat loss rate is therefore <1%.

3.8 THE AEROBIC REACTOR WALL HEAT LOSSES

3.8.1 Introduction

Energy is continually lost from the system by the transfer of heat through the walls of the aerobic reactor. This transfer of energy takes place by virtue of the fact that the temperature of the materials within the reactor (the sludge, foam and effluent gas) are at a higher temperature than the materials outside the reactor (the anaerobic sludge, digester biogas and atmosphere).

This heat loss H_w, which describes the total wall heat loss from the aerobic reactor is a relatively minor term in the heat balance, contributing less than 10 percent to the total heat loss from the system. It is one of the benefits of the design of the Athlone Dual Digester that the placement of the aerobic reactor within the anaerobic digester (and not as a separate unit) improves the insulation of the reactor, thereby reducing the wall heat losses.

The overall wall heat loss from the aerobic reactor H_w is calculated from an application of heat transfer theory. It would have been possible to perform an experimental assessment of the wall heat loss by stopping the feed sludge and air supply to the reactor. Under such conditions, at an equilibrium temperature, the rate of heat loss through the walls of the reactor would be equal to the rate of mechanical heat input from the mixing device $H_m = H_w$ (with all other terms in the Heat Balance being zero).

The decision against performing an experimental assessment in this manner was made for the following three reasons: (1) It was considered that stopping the feed sludge and air supply would have had a detrimental effect on the population of thermophilic bacteria; (2) the determination of the mechanical heat input from the mixing device is itself obtained by estimation (see Section 3.9 below); and (3) the assessment would have excluded any heat loss from the foam and gas in the reactor head space. As the wall heat loss is a relatively minor term in the heat balance it was considered justifiable to carry out only a theoretical estimation.

3.8.2 Requirements for a Theoretical Estimation of the Wall Heat Losses

In order to perform a theoretical estimation of the wall heat losses from the aerobic reactor, consideration needs to be given to the mechanism by which heat transfer takes place at any given point around the reactor wall. When a temperature difference exists between 2 parts of a system the transfer of heat will result. This heat transfer can take place in one or more of 3 different ways: by conduction, convection or radiation.

The magnitude of the heat flow at any given point, and the predominance for either conduction, convection or radiation as the mechanism for heat transfer, is dependent upon: (1) the thermal properties of the materials through which the heat must flow, (2) the fluid dynamics of any liquids or gases involved, (3) the geometry of the path of heat transfer, and (4) the temperature difference which exists between the two points in the system.

In order for the overall heat loss from the aerobic reactor to be determined, it is necessary that: (1) the mechanism by which heat transfer takes place at each point around the reactor is established, (2) the different conditions which exist both inside and outside the reactor wall are identified, (3) the thermal properties of the materials concerned are known, (4) the fluid dynamics of the different sludge and gas regimes are assessed, (5) appropriate temperature measurements are taken both inside and outside the reactor which can adequately describe the temperature gradients which exist around the system, and (6) mathematical expressions are developed which quantify the heat which is transferred at any given point.

3.8.3 Identification of the Different Mechanisms of Heat Transfer

A review of the plan of the dual digester (Figure 3.10) reveals six distinct areas (labelled **A** to **F**) here conditions either side of the reactor wall differ from one another.

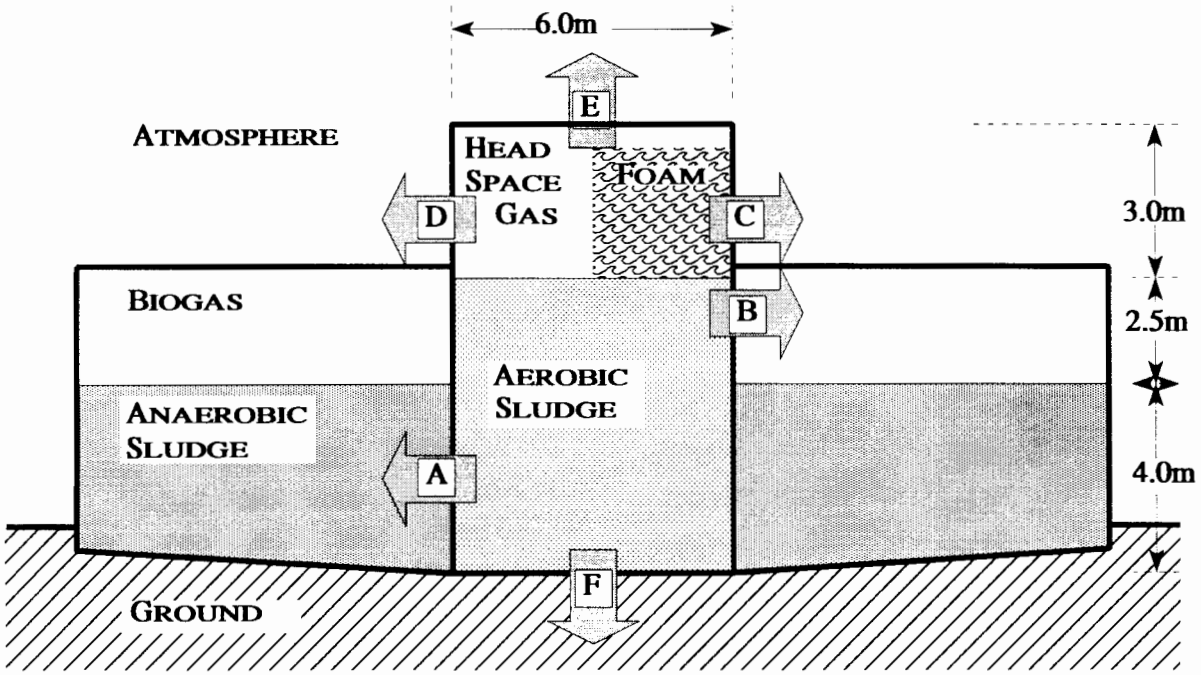

Figure 3.10 Identification of the Different Areas of Wall Heat Loss

Heat within the aerobic reactor is lost through the walls from: the aerobic sludge (**A**, **B** and **F**), the foam layer if present (**C**), and the gas in the reactor head space (**D** and **E**). The six areas are defined in Table 3.7.

Table 3.7 Description of the Six Areas of Wall Heat Loss

Area	Description of Area	m^2
A	The Vertical Walls in the Bottom 4.0m Section of the Reactor	80.4
B	The Vertical Wall in the Centre 2.5m Section of the Reactor	50.3
C	The Vertical Walls in the Top 3.0m Section of the Reactor	60.3
D	The Vertical Walls in the Top 3.0m Section of the Reactor	60.3
E	The Roof of the Reactor	28.3
F	The Base of the Reactor	28.3

The bottom 4.0m section of the reactor can be considered to be immersed in a bath of anaerobic sludge, the middle 2.5m section insulated by the digester biogas, and the top 3.0m section exposed to the atmosphere. The base of the reactor is in direct contact with the ground. The different mechanisms by which heat transfer takes place from the aerobic reactor are as follows:

(1) Heat is transferred from the aerobic sludge and foam layer to the inner surface of the reactor wall by **conduction and forced convection**.

(2) Heat is transferred from the saturated gas stream to the inner surface of the reactor walls in the gas head space by **conduction**.

(3) Heat is transferred through the concrete walls of the reactor by **conduction**.

(4) Heat is transferred from the outer surface of the reactor wall to the anaerobic sludge by **conduction and forced convection**.

(5) Heat is transferred from the outer surface of the reactor wall to both the digester biogas and atmosphere by **natural convection and radiation**.

As in most applications of this nature, any heat loss through the base of a system (in this instance Area F) in contact with the ground is considered negligible or non-existent on the basis that: if one goes deep enough then the temperature of the ground will start to increase and thus eliminate any temperature gradient between the ground and the system.

3.8.4 Heat Transfer Theory

Conduction

Conduction is the transfer of heat from one part of a body to another part of the same body, or from one body to another in physical contact with it. For non-metallic solids (i.e. concrete), conduction is the result of the transfer of vibrational energy from one molecule to another within the solid. The fundamental differential equation for describing heat transfer by conduction is derived from Fourier's Law. For the flow of heat through a plane wall of thickness x Fourier's Law takes the form:

$$Q = \frac{kA(T(w)_{in} - T(w)_{out})}{x} \quad \text{...MJ/h} \quad (3.164)$$

where:

Q = Rate of heat transfer by conduction (MJ/h)
k = Thermal conductivity which is temperature dependent (MJ/h.m.°C)
A = Area at right angles to the direction of heat flow (m^2)
$T(w)_{in}$ = Inside surface temperature of the wall (°C)
$T(w)_{out}$ = Outside surface temperature of the wall (°C)
x = Thickness of the wall (m)

From Perry and Chilton (1973) the thermal conductivity of concrete is quoted as 0.0027 MJ/h.m.°C at a temperature of 20°C. Accepting that the thermal conductivity does not change significantly over the temperature range encountered in the aerobic reactor, then this constant value for k can be accepted. The side walls of the aerobic reactor are a uniform 0.40m thickness. The roof thickness was actually slightly less than the sidewalls and contained a large opening for inspection purposes. A thick rubber mat then covered the roof, increasing its insulation. For simplicity the thickness of the roof is taken to be the same as that of the sidewalls. From Eq 3.164 the flow of heat by conduction through a plane concrete wall of thickness 0.40m is as follows:

$$Q = 0.00675 \times A(T(w)_{in} - T(w)_{out}) \quad \text{...MJ/h} \quad (3.165)$$

Forced Convection

Forced Convection is the transfer of heat from one point to another within a fluid (gas or liquid) by the enforced mixing (i.e. through the use of a mixing pump) of one portion of the fluid with another.

In most cases where convective heat transfer is taking place between a fluid and the surface of a wall (or vice versa), the circulating currents die out in the immediate vicinity of the surface and a thin film of fluid, free of turbulence, covers the surface of the wall. The transfer of heat through the thin film takes place by conduction. When the body of the fluid is well mixed, the resistance to the transfer of heat between the fluid and the surface of the wall can be regarded as lying within the thin film covering the surface.

If it is assumed that the resistance to the transfer of heat through the liquid film(s) at the surface of the walls is negligible in comparison to the resistance encountered within the walls, and that there is negligible scale build-up on the walls, then the surface temperature of the wall can be assumed to be the same as the bulk liquid. Indeed if the

mixing of the bulk liquid is efficient, as is the case for the aerobic reactor (and to a lesser extent for the anaerobic digester) this is an entirely reasonable assumption to make. It is therefore unnecessary for surface temperature measurements to be taken in the areas where there is liquid contact with the surface of the wall.

Natural Convection and Radiation

Heat transfer from the surface of a wall to air (or gas) predominantly takes place by both natural convection and radiation. Natural convection occurs as a result of (1) the mixing action brought on by differences in gas density arising from temperature gradients in the gas and (2) the enforced mixing as a result of turbulent gas conditions at and around the surface of the wall. Heat transfer by radiation takes place in the form of electromagnetic wave motion through space. All materials radiate thermal energy in this manner if they are at a higher temperature than their surroundings.

The rate of heat transfer by convection from the hot surface of a wall through an air/gas medium to its cooler surroundings, accepting the temperature of the surroundings to be at the air/gas temperature[1] $T(GAS)$, is given by:

$$Q = h_c A \left(T(w)_{out} - T(GAS) \right) \quad \text{...MJ/h} \quad (3.166)$$

where:

h_c = Heat transfer coefficient for convection (MJ/m^2.h.°C)

$T(GAS)$ = Temperature of the surrounding air/gas (°C)

Expressions for the heat transfer coefficient for convection have been developed very much on an empirical basis and are dependent on the geometry of the system and the fluid dynamics of the gas. For the turbulent motion of air/gas across a plane *vertical* wall the heat transfer coefficient is given by Perry and Chilton (1973) as $h_c = 0.0072\ \Delta T^{0.25}$. Under such conditions the rate of heat transfer by convection is therefore as follows:

$$Q = 0.0072A \left(T(w)_{out} - T(GAS) \right)^{1.25} \quad \text{...MJ/h} \quad (3.167)$$

For the turbulent motion of air/gas above a *horizontal* plane wall the heat transfer coefficient is given by Perry and Chilton (1973) as $h_c = 0.0084\Delta T^{0.25}$. Under such conditions the rate of heat transfer by convection is therefore as follows:

$$Q = 0.0084A \left(T(w)_{out} - T(GAS) \right)^{1.25} \quad \text{...MJ/h} \quad (3.168)$$

---

[1] The term $T(GAS)$ is used in the derivation, referring to both the temperature of the biogas in the digester headspace and the temperature of the outside air.

The rate of heat transfer by **radiation** from the surface of a wall through a gas medium to its surroundings, again accepting the temperature of the surroundings to be at the air/gas temperature is given by:

$$Q = h_r A\left((273 + T(w)_{out}) - (273 + T(GAS)) \right) \quad \text{...MJ/h} \quad (3.169)$$

where:

h_r = Heat transfer coefficient for radiation (MJ/m^2.h.K)

An expression for the heat transfer coefficient for radiation is derived from an application of both Kirchoff's Law and the Stefan-Boltzman Law and is as follows:

$$h_r = \sigma e \frac{\left((273 + T(w)_{out})^4 - (273 + T(GAS))^4 \right)}{\left((273 + T(w)_{out}) - (273 + T(GAS)) \right)} \quad \text{...MJ/m}^2\text{.h.K} \quad (3.170)$$

where:

σ = Stefan-Boltzmann constant (MJ/m^2.h.K^4)

e = Emissivity of the grey body

The Stefan-Boltzmann constant σ is equal to 2.04 x 10^{-10} MJ/m^2.h.K^4. The emissivity e of concrete as quoted by Perry and Chilton (1973) is 0.90. Substitution into Eq 3.170 yields:

$$h_r = 1.84 \times 10^{-10} \frac{\left((273 + T(w)_{out})^4 - (273 + T(GAS))^4 \right)}{\left((273 + T(w)_{out}) - (273 + T(GAS)) \right)} \quad \text{...MJ/m}^2\text{.h.K} \quad (3.171)$$

Substitution of Eq 3.171 for h_r into Eq 3.169 for the rate of heat transfer by radiation from the surface of a wall to its surroundings, yields:

$$Q = 1.84 \times 10^{-10} A\left((273 + T(w)_{out})^4 - (273 + T(GAS))^4 \right) \quad \text{...MJ/h} \quad (3.172)$$

The total heat transferred by the combination of natural **convection and radiation** from the surface of a wall through an air/gas medium to the surrounds can be obtained firstly by combining the fundamental equations Eq 3.166 and Eq 3.169:

$$Q = A(h_c + h_r)\left(T(w)_{out} - T(GAS) \right) \quad \text{...MJ/h} \quad (3.173)$$

and then, substituting for the heat transfer coefficients h_c and h_r i.e.

for a *vertical* wall

$$Q = 0.0072A(T(w)_{out} - T(GAS))^{1.25} + 1.84 \times 10^{-10} A((273 + T(w)_{out})^4 - (273 + T(GAS))^4) \quad \text{MJ/h} \quad (3.174)$$

and for a *horizontal* wall

$$Q = 0.0084A\left(T(w)_{out} - T(GAS)\right)^{1.25} + 1.84\times10^{-10}A\left((273 + T(w)_{out})^4 - (273 + T(GAS))^4\right) \quad \text{MJ/h} \quad (3.175)$$

Conduction through a film of Condensate

When a saturated vapour is brought into contact with a cold surface, heat is transferred from the vapour to the surface and condensation on the surface of the wall takes place.

For a vapour condensing on a vertical surface, a film of condensate is produced which flows downwards under the influence of gravity and is retarded by the viscosity of the liquid itself. The flow will normally be streamline and heat will flow through the film by conduction. It can be assumed that the resistance to the transfer of heat through the liquid film is negligible in comparison to the resistance encountered within the walls. The surface temperature of the vertical walls in the aerobic reactor head space can therefore be assumed to be at the same temperature as the gas in the reactor head space.

When a vapour condenses on the underside of a horizontal surface, droplets are formed on the underside of the roof, which do not wet the surface, but after growing slightly they will fall from the roof exposing fresh condensing surface. This gives what is known as drop wise condensation and, since the heat does not have to flow through the film by conduction, much higher transfer coefficients are obtained. It is accepted that, under such conditions, the temperature of the inner surface of the aerobic reactor roof is at the same temperature as that of the gas in the reactor headspace.

It is appropriate at this stage to consider the mass transfer of water vapour from the centre of the reactor gas head space to the side walls in order for condensation to take place. If the velocity of the gas stream is too high, i.e. with a short retention time in the reactor head space, then the condensation effects described above will be greatly reduced. At an air flow rate of 760 m^3/h the head space gas retention time is calculated at approximately 11 minutes. This is considered adequate time for completely mixed conditions to be achieved in the head space and for the gas stream velocity not to inhibit condensation. This fact was confirmed experimentally by the even temperature profile observed around the reactor head space on the occasions when the gas stream temperature was measured.

3.8.5 The Required Temperature Measurements to Monitor Wall Heat Loss

The temperature measurements which were taken to monitor the heat losses from the aerobic reactor were as follows:

$T(AIR)_{in}$ = Influent gas stream temperature (°C)
$T(AIR)_{out}$ = Effluent gas stream temperature (°C)
$T(SL)_{in}$ = Influent feed sludge temperature (°C)
$T(SL)_r$ = Reactor sludge temperature (°C)
$T(SL)_d$ = Digester sludge temperature (°C)

The following five assumptions are made in order to simplify the eventual estimate of the wall heat loss:

(1) The atmospheric temperature, and that of the surroundings, are accepted to be at the same temperature as the influent gas stream to the reactor $T(AIR)_{in}$ which was measured daily (see Table 2.4).

(2) The temperature of the foam layer is accepted to be at the same temperature of the sludge in the reactor $T(SL)_r$. On the occasions when the foam temperature was measured (once or twice during foam periods) this was found to be the case.

(3) In spite of the fact that the aerobic reactor gas stream undergoes some degree of cooling in the reactor head space, for ease of calculation and to obviate the need to distinguish between foaming and nonfoaming conditions when setting up an expression for the wall heat loss, the temperature of the gas stream in the aerobic reactor head space is accepted to be at the same temperature as that of the sludge in the reactor $T(SL)_r$.

Whilst the drop in the effluent gas stream temperature below that of the sludge is significant when calculating the vapour heat loss from the system (see Section 3.5), for the purposes of estimating the wall heat loss in this region, it is considered that an exact solution in terms of the reactor head space gas temperature is not warranted.

(4) The temperature of the biogas in the digester head space is accepted to be at the same temperature as the sludge in the digester $T(SL)_r$. Undoubtably the biogas is significantly cooled by contact with the cooler outside walls of the digester, probably far more so than the gas stream in the reactor. Again, a precise solution

for the wall heat loss in terms of biogas temperature (reduced through cooling) is not warranted.

(5) In order to quantify the heat which is transferred away from the reactor walls by natural convection and radiation (Areas **B**, **C**, **D** and **E**) knowledge of the outside wall surface temperature for each area is required (Eq 3.174 and Eq 3.175). The wall temperature $T(w)_{out}$ is determined theoretically by performing an energy balance at the surface of the wall on the principal that the quantity of heat which is transferred away from the wall by convection and radiation must be equal to the quantity of heat which is conducted through the wall.

In line with assumption (5) above, equating the rate at which heat is transferred away from the wall by convection and radiation (Eq. 3.173) with the rate at which heat is conducted through the wall (Eq. 3.164) yields:

$$Q = \frac{kA(T_1 - T_2)}{x} = A(h_c + h_r)(T_2 - T_3) \quad \text{...MJ/h} \quad (3.176)$$

where

Q = Heat flow rate (MJ/h)

T_1 = Temperature of substance/medium inside the reactor (°C)

T_2 = Temperature of the outside wall surface (°C)

T_3 = Temperature of substance/medium outside the reactor (°C)

Both T_1 and T_3 are measurable parameters, either taken from temperature readings made directly, or deduced in terms of assumptions (1) to (4). The unknown parameter, the outside wall surface temperature, T_2 can be calculated by iterating once T_1 and T_3 are known (the heat transfer coefficients h_c and h_r are both complex temperature dependent functions which makes it impossible to derive a direct mathematical solution for T_2 in terms of T_1 and T_3). As it is inconvenient to perform an iteration for each set of data, the iteration is performed once under a given set of conditions. The temperature difference $(T_1 - T_2)$ across the wall is then expressed as a fixed fraction f_Δ of the temperature difference between the materials/medium inside and outside the reactor walls $(T_1 - T_3)$ viz:

$$(T_1 - T_2) = f_\Delta(T_2 - T_3) \quad \text{...°C} \quad (3.177)$$

Once the fraction f_Δ has been established by iteration for a particular set of conditions, the heat transferred through the wall and then ultimately away from the wall by

convection and radiation, for all other temperatures T_1 and T_3 can be adequately estimated by:

$$Q = f_\Delta 0.00675 A (T_1 - T_3) \quad \text{...MJ/h} \quad (3.178)$$

3.8.6 Application of Heat Transfer Theory

Determination of the Heat Loss from each Individual Area

Case A describes the heat transfer through the side wall area (80.4m^2) in the bottom section of the reactor. The rate of heat transfer from the hot aerobic sludge, by conduction (Eq 3.103) through the reactor wall, to the cooler anaerobic sludge is as follows:

$$H_w^a = 0.00675 \times 80{,}4 (T(SL)_r - T(SL)_d) = 0.54 (T(SL)_r - T(SL)_d) \quad \text{...MJ/h} \quad (3.179)$$

Case B describes the heat transfer through the side wall area (50.3m^2) in the centre section of the reactor. Heat is transferred from the hot aerobic sludge, by conduction (Eq 3.103) through the reactor wall. This heat is then transferred by radiation and convection (Eq 3.113) into the digester head space. The energy balance at the outside wall surface (from Eq 3.115) is as follows:

$$\begin{aligned} 0.00675 \times 50.3 (T(SL)_r - T(w)_{out}) &= 0.0072 \times 50.3 (T(w)_{out} - T(SL)_d)^{1.25} \\ &+ 1.84 \times 10^{-10} \times 50.3 ((273 + T(w)_{out})^4 - (273 + T(SL)_d)^4) \end{aligned} \quad \text{...MJ/h} \quad (3.180)$$

For $T(SL)_r$ = 50°C and $T(SL)_d$ = 30°C, a value of $T(w)_{out}$ = 33.6°C is obtained from the above equation by iteration. It follows therefore that:

$$f_\Delta = \frac{(50 - 33.6)}{(50 - 30)} = 0.82 \quad \text{...} \quad (3.181)$$

For all conditions therefore, the temperature drop across the wall is given by:

$$(T(SL)_r - T(w)_{out}) = 0.82 \times (T(SL)_r - T(SL)_d) \quad \text{...°C} \quad (3.182)$$

The heat transferred by conduction through the wall in area B (from Eq 3.178) is as follows:

$$H_w^b = 0.82 \times 0.00675 \times 50.3 (T(SL)_r - T(SL)_d) = 0.28 (T(SL)_r - T(SL)_d) \quad \text{...MJ/h} \quad (3.183)$$

Case C describes the heat transfer through the side wall area (60.3m^2) in the upper section of the reactor when foam is present on the surface of the sludge. Heat is transferred from the hot aerobic foam, by conduction (Eq 3.164) through the reactor wall. This heat is then transferred by radiation and convection (Eq 3.174) to the surrounding

environment. The energy balance at the outside wall surface (from Eq 3.176) is as follows:

$$0.00675 \times 60.3(T(SL)_r - T(w)_{out}) = 0.0072 \times 60.3(T(w)_{out} - T(AIR)_{in})^{1.25} + 1.84 \times 10^{-10} \times 60.3((273 + T(w)_{out})^4 - (273 + T(AIR)_{in})^4) \quad \text{...MJ/h} \quad (3.184)$$

For $T(SL)_r$ = 50°C and $T(AIR)_{in}$ = 20°C, a value of $T(w)_{out}$ = 25.5°C is obtained from the above equation by iteration. It follows therefore that:

$$f_\Delta = \frac{(50 - 25.5)}{(50 - 20)} = 0.82 \quad \text{...} \quad (3.185)$$

For all conditions therefore, the temperature drop across the wall is given by:

$$(T(SL)_r - T(w)_{out}) = 0.82 \times (T(SL)_r - T(AIR)_{in}) \quad \text{...°C} \quad (3.186)$$

The heat transferred by conduction through the wall in area C (from Eq 3.178) is as follows:

$$H_w^b = 0.33(T(SL)_r - T(AIR)_{in}) \quad \text{...MJ/h} \quad (3.187)$$

Case D describes the heat loss from the upper section of the reactor when no foam is present on the surface of the sludge. Heat is transferred from the effluent gas stream to the inside walls of the reactor head space as a result of condensation. This heat is then conducted (Eq 3.164) through the walls of the reactor and transferred by radiation and convection (Eq 3.174) to the surrounding environment. As the inside wall temperature is considered to be fixed by the reactor sludge temperature, it is therefore independent of whether foam is present on the surface of the sludge or not. The heat which is transferred from the upper section side walls under non-foaming conditions is the same as when foaming occurs. It therefore follows that (from Eq 3.187):

$$H_w^d = H_w^c = 0.33(T(SL)_r - T(AIR)_{in}) \quad \text{...MJ/h} \quad (3.188)$$

Case E describes the heat transfer through the roof of the reactor (28.3m^2). Heat is transferred from the effluent gas stream to the underside of the reactor roof as a result of condensation. This heat is then conducted (Eq 3.164) through the roof wall and transferred by radiation and convection (Eq 3.174) to the surrounding environment. The energy balance at the outside surface of the roof (from Eq 3.176) is as follows:

$$0.00675 \times 28.3(T(SL)_r - T(w)_{out}) = 0.0084 \times 28.3(T(w)_{out} - T(AIR)_{in})^{1.25} + 1.84 \times 10^{-10} \times 28.3((273 + T(w)_{out})^4 - (273 + T(AIR)_{in})^4) \quad \text{...MJ/h} \quad (3.189)$$

For $T(SL)_r$ = 50°C and $T(AIR)_{in}$ = 20°C, a value of $T(w)_{out}$ = 25.3°C is obtained from the above equation by iteration. It follows therefore that:

$$f_\Delta = \frac{(50 - 25.3)}{(50 - 20)} = 0.82 \quad \ldots (3.190)$$

For all conditions therefore, the temperature drop across the roof wall is given by:

$$(T(SL)_r - T(w)_{out}) = 0.82 \times (T(SL)_r - T(AIR)_{in}) \quad \ldots °C \quad (3.191)$$

The heat transferred by conduction through the wall in area E (from Eq 3.187) is as follows:

$$H_w^e = 0.82 \times 0.00675 \times 28.3 (T(SL)_r - T(AIR)_{in}) = 0.28 (T(SL)_r \quad \ldots MJ/h \quad (3.192)$$

Case F describes the heat loss through the base of the system, which as discussed in Section 3.8.3 is considered to be negligible and therefore ignored in the calculation of the overall wall heat loss from the reactor (i.e. $H_w^f \approx 0$).

3.8.7 Calculation of the Overall Wall Heat Losses

The overall wall heat loss from the aerobic reactor is obtained by summing the individual heat loss terms described above. Note that Case C and D are identical in magnitude and therefore no distinction is made between foaming or non foaming conditions and case F (the heat loss through the bottom of the aerobic reactor) is accepted to be zero. Hence:

$$H_w = H_w^a + H_w^b + H_w^c (H_w^d) + H_w^e \quad \ldots MJ/h \quad (3.193)$$

Substitution of the appropriate formulae yields:

$$H_w = 0.54 (T(SL)_r - T(SL)_d) + 0.28 (T(SL)_r - T(SL)_d) + 0.33 (T(SL)_r - T(AIR)_{in}) + 0.16 (T(SL)_r - T(AIR)_{in}) \quad \ldots MJ/h \quad (3.194)$$

which yields upon simplification:

$$H_w = 0.82 (T(SL)_r - T(SL)_d) + 0.49 (T(SL)_r - T(AIR)_{in})$$

Equation 3.195 The Wall Heat Loss Rate H_w (MJ/h)

A graphical representation of Equation 3.195 for the rate of wall heat loss from the reactor H_w with reactor sludge temperature $T(SL)_r$ is given below in Figure 3.11.

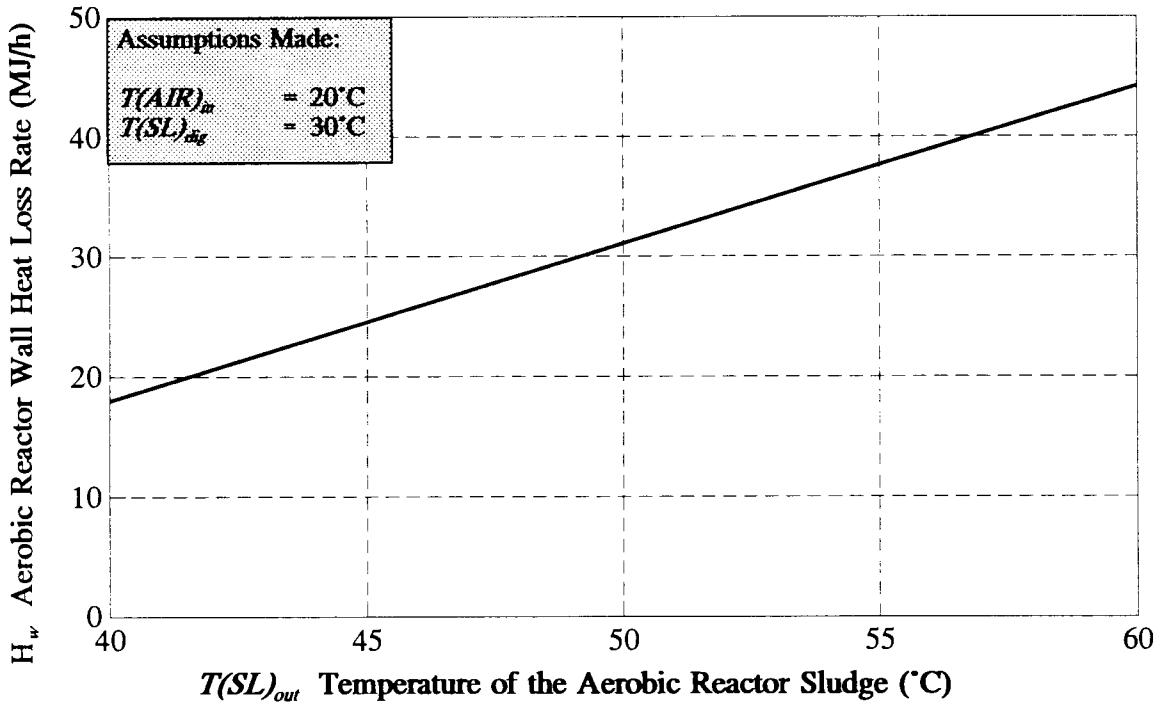

Figure 3.11 The Aerobic Reactor Wall Heat Loss Rate H_w versus Reactor Sludge Temperature $T(SL)_{out}$ from Equation 3.195. The Influent Air and Digester Sludge Temperatures are taken to be 20°C and 30°C respectively.

Comparing Figure 3.11 for the wall heat loss rate H_w with Figure 3.8 for the vent gas sensible heat loss rate H_g , it can be seen that at an air flow rate of 760m^3(STP)/h H_w is of a similar magnitude as H_g , i.e. each between 20 and 40 MJ/h which for each is less than about 6% of the total heat losses. In Section 3.10.5 below, a summary view of all the heat losses and gains from the reactor is provided, both for oxygenation with air alone and air plus pure oxygen.

3.9 MECHANICAL HEAT GENERATION

3.9.1 The Benefits of Efficient Mixing

For the aerobic reactor process to work effectively, it is essential that the contents of the reactor are **continuously** and **thoroughly** mixed. Efficient mixing promotes intimate contact between the thermophilic bacteria, the dissolved oxygen, and the organic substrate maximising the biological activity in the reactor. The aerobic process is further benefited by mixing, in that the kinetic energy generated by the mechanical action of the

mixing device is ultimately transferred to **thermal energy** through collisions which take place between the sludge particles, and is therefore a source of heat to the system. The rate at which this transfer of heat takes place can be estimated from the power consumption of the mixing device by making allowances for the inefficient transfer of energy by the pump motor and some heat transmission losses in the bearings and couplings.

3.9.2 Method of Mixing in the Athlone Aerobic Reactor

The contents of the aerobic reactor at Athlone were mixed by pumping the sludge through a recirculation line (see Figure 2.3). During phase I, when oxygenation was with air alone, the only criterion with respect to mixing was to provide a turnover of reactor contents every half-hour. This was achieved by operating two identical centrifugal pumps, each fitted with a 7.5kW motor, alternately at fixed time intervals (see Section 2.3.6 for further details).

During phase II pure oxygen was injected into the sludge recirculation line, immediately downstream of the pumps, to supplement the aeration system. Due to the requirements for successful oxygen injection and dissolution, the recirculation line was modified to allow for increased pressure at the point of injection and increased velocity at discharge (see Section 2.3.7 for details). At the start of phase II (days 1-25), the sludge was circulated using one centrifugal pump fitted with a 55kW motor. Between days 26 and 40, the sludge was circulated using the second pump fitted with a 75kW motor. Due to the poor oxygen transfer efficiency obtained during both periods (see Section 7.3.3 for details) the pipework was modified to allow for both pumps to run together in series for the remainder of phase II (days 74-152).

3.9.3 Power Consumption of the Mixing Device

The electrical power consumed in a three-phase balanced Y system with phase voltage V and phase current drawn I is given by the following formula, extracted from Perry and Chilton (1973):

$$P = \sqrt{3} \times V \times I \times \cos\phi \quad \text{...kW} \quad (3.196)$$

where:

P = Electrical power drawn (kW)
V = Phase voltage (kV)
I = Phase current drawn (amps)
$\cos\phi$ = Power factor [2]

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[2] The power factor $\cos\phi$ is the ratio of the power consumption to the volt-amperes

The phase current (I) drawn by the electric motor(s) on each pump was measured daily from the pump ammeter. The phase voltage (V) was 380 volts. A value of 0.82 was accepted for the power factor cosϕ, estimated from engineering data for the operational conditions of each motor. Substitution of this data into Eq 3.196 yields:

$$P = \sqrt{3} \times 0.380 \times I \times 0.82 = 0.54 \times I \quad \text{...kW} \quad (3.197)$$

3.9.4 Conversion of Electrical Power to Mechanical Heat

The energy loss due to (1) the inefficient transfer of energy from the pump motor and (2) the heat transmission losses in the bearings and couplings can be assumed to remain constant. The mechanical heat generated by the mixing pumps can therefore be related to the electrical power drawn by the motors via the constant of proportionality f_{mech} viz:

$$H_m = f_{mech} \times P \times 3.6 \quad \text{...MJ/h} \quad (3.198)$$

where:

H_m = Mechanical heat generation rate (MJ/h)

f_{mech} = Fraction of electrical power drawn by the pump motor which is converted to heat in the reactor sludge

P = Electrical power drawn by the pump (kW)

3.6 = Conversion factor kW to MJ/h

The total energy loss in the conversion of electrical power to mechanical heat is not normally more than 10-15% (i.e. $1.0 > f_{mech} > 0.85$). On the Milnerton pure oxygen aerobic reactor, the fraction of power converted to heat (f_{mech}) was measured experimentally and found to be 0.85 (Messenger *et al*, 1992). Because of the similarity between the mixing device at Milnerton and that at Athlone, it was considered appropriate to use the same value for f_{mech} to estimate the mechanical heating rate at Athlone.

3.9.5 Calculation of the Mechanical Heating Rate

Substituting Eq 3.197 for the power consumption P into Eq 3.198 for the Mechanical Heating Rate H_m and accepting the value of 0.85 for the energy transfer efficiency factor f_{mech} yields:

$$H_m = 0.85 \times 0.54 \times I \times 3.6 \quad \text{...MJ/h} \quad (3.199)$$

which upon simplification yields:

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and hence must be equal to or less than 1.0. The volt-amperes in an alternating current circuit is frequently greater than the power consumption due to the effect of inductance or capacity. ϕ is the angular phase difference between voltage and current.

$$H_m = 1.65 \times I$$

Equation 3.200 The Mechanical Heat Generation Rate H_m (MJ/h)

3.10 BIOLOGICAL HEAT GENERATION

3.10.1 Introduction

The importance of biological heat generation in autothermal aerobic processes has been fully discussed in Section 3.1 above. The rate of biological heat generation in the thermophilic aerobic reactor of the dual digestion system has been directly linked to the rate of oxygen utilisation via the specific heat yield coefficient viz:

$$H_b = Y_h . OUR . V_p \quad \text{...MJ/h} \quad (3.201)$$

where:

H_b = Biological heat generation rate (MJ/h)
Y_h = Specific heat yield (MJ/kg(O_2))
V_p = Process volume (m^3)
OUR = Oxygen utilisation rate (kg(O_2)/m^3.h)

If a value for the specific heat yield is known or assumed, then the biological heat generated can be calculated directly by measuring the oxygen utilisation rate. For example, by accepting the value for the specific heat yield of 13.0 MJ/kg(O_2) measured by Messenger *et al* (1992), the biological heat generation rate is given by:

$$H_b = 13.0 \times OUR . V_p \quad \text{...MJ/h} \quad (3.202)$$

However one of the objectives of this investigation was to determine the specific heat yield for conditions in the Athlone aerobic reactor. Rearranging Eq 3.201, the specific heat yield is as follows:

$$Y_h = \frac{H_b}{OUR . V_p} \quad \text{...MJ/kg}(O_2) \quad (3.203)$$

It is apparent therefore, that in order to determine the specific heat yield, both the **biological heating rate** and the **oxygen utilisation rate** need to be determined under a variety of different operating conditions, such as retention time and influent air flow rate.

3.10.2 Determination of the Biological Heating Rate from the Heat Balance

The quantity of heat which is generated biologically within the aerobic reactor is determined from rearranging the steady state heat balance (Eq. 3.8) in terms of H_b as follows:

$$H_b = H_g + H_s + H_v + H_w - H_m \quad \text{...MJ/h} \quad (3.204)$$

where:

H_s = The rate of sensible heat loss with the sludge leaving the reactor
H_v = The rate of vapour heat loss with the effluent gas stream
H_g = The rate of sensible heat loss with the effluent gas stream
H_w = The rate of heat loss through the walls of the reactor
H_m = The rate of heat energy input from the mixing device

Formulae to allow for the estimation of each of the terms on the right hand side of Equation 3.204 above have been developed, both for oxygenation with air alone and for oxygenation with air and pure oxygen, in terms of readily measurable parameters. The formulae applicable when oxygenation is with air alone (phase I) are as follows:

The Water Vapour Heat Loss Rate in the Effluent Gas (MJ/h) (Eq 3.129)
(Valid for Oxygenation with Air Alone)

$$H_v = 2.38 \times 1.205\, Q(AIR)_{in} \left(\frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{0.62 \times alog_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)}{1240 - alog_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)} \right)$$

The Effluent Gas Sensible Heat Loss Rate (MJ/h) (Eq 3.151)
(Valid for Oxygenation with Air Alone)

$$H_g = 1.21\, Q(AIR)_{in} \left(0.001\,(T(AIR)_{out} - T(AIR)_{in}) + \frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} 0.00187\, T(AIR)_{out} \right)$$

The Sludge Sensible Heat Loss Rate (MJ/h) (Eq 3.161)
(Valid for Oxygenation with Air Alone)

$$H_s = 4.04 \left(\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{1.21\, T(SL)_r Q(AIR)_{in}}{1000} \times \frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 - T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 - T(AIR)_{out}}\right)} - \frac{Q(SL)_{in} T(SL)_{in}}{24} \right)$$

The Rate of Heat Loss from the Walls of the Reactor (MJ/h) (Eq 3.195)
(Valid for Oxygenation with Air Alone and with Air + Pure Oxygen)

$$H_w = 0.82\left(T(SL)_r - T(SL)_d\right) + 0.49\left(T(SL)_r - T(AIR)_{in}\right)$$

The Rate of Mechanical Heat Input to the Reactor (MJ/h) (Eq 3.200)
(Valid for Oxygenation with Air Alone and with Air + Pure Oxygen)

$$H_m = 1.65 \times I$$

The measurable parameters contained in the above formulae are as follows:

$Q(AIR)_{in}$ = The dry air volumetric flow rate into the reactor (m^3(STP)/h)
$T(AIR)_{out}$ = The influent gas stream temperature to the aerobic reactor (°C)
$T(AIR)_{out}$ = The effluent gas stream temperature from the aerobic reactor (°C)
$T(SL)_r$ = The temperature of the sludge leaving the reactor (°C)
$T(SL)_{in}$ = The temperature of the feed sludge to the reactor (°C)
$Q(SL)_{in}$ = The feed sludge flow rate (m^3/d)
$T(SL)_d$ = The temperature of the sludge in the digester (°C)
I = The current drawn by the mixing pump (amperes)

By operating the aerobic reactor under steady state conditions and monitoring the above parameters (phase I), each of the heat loss and gain terms in the steady state heat balance were determined and consequently the rate of biological heat generation was calculated by difference. The computer programme written to perform the necessary computation is given in Appendix 8.

During phase II the aerobic reactor was oxygenated with air and pure oxygen. Due to the increased gas flow through the system the vapour heat loss, the gas sensible heat loss, and the sludge sensible heat loss rates changed. Formulae to allow for the estimation of these heat loss terms during pure oxygen supplementation were developed above, and are as follows:

The Water Vapour Heat Loss Rate in the Effluent Gas (MJ/h) (Eq 3.133)
(Valid for Oygenation with Air + Pure Oxygen)

$$H_v^{AIR+O2} = 2.38\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right)\cdot\left(\frac{0.62a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{0.62a\log_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)}{1240 - a\log_{10}\left(9.12 - \frac{2307}{273 + T(AIR)_{in}}\right)}\right)$$

The Effluent Gas Sensible Heat Loss Rate (MJ/h) (Eq 3.153)
(Valid for Oygenation with Air + Pure Oxygen)

$$H_g^{AIR+O2} = \left(1.21Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left[0.001\left(T(AIR)_{out} - T(AIR)_{in}\right) + \frac{0.62a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} 0.00187T(AIR)_{out}\right]$$

The Sludge Sensible Heat Loss Rate (MJ/h) (Eq 3.163)
(Valid for Oygenation with Air + Pure Oxygen)

$$H_s^{AIR+O2} = 4.04\left[\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{T(SL)_r\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right)}{1000}.\frac{0.62a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{Q(SL)_{in}T(SL)_{in}}{24}\right]$$

The additional measurable parameter now included to allow solution of the above is the pure oxygen mass flow rate into the reactor $M(O_2)_{in}^{O2}$ (see Section 2.4.7).

3.10.3 Determination of the Specific Heat Yield Coefficient

Once the biological heating rate is calculated from the steady state heat balance, the specific heat yield coefficient can be determined from the calculated oxygen utilisation rate *(**OUR**)*. i.e.

$$Y_h = \frac{H_g + H_s + H_v + H_w - H_m}{OUR.V_p} \quad \text{...MJ/kg}(O_2) \quad (3.205)$$

In Section 3.2 an equation (Eq. 3.44) was derived for calculating the oxygen utilisation rate, for oxygenation using air alone, from the oxygen gas mass balance across the reactor viz:[3]

$$OUR = \frac{Q(AIR)_{in}}{138}.\frac{\left(21 - \%(O_2)_{out}\right)}{\left(100 - 0.3 \times \%(O_2)_{out}\right)} \quad \text{...kg}(O_2)/\text{m}^3\text{.h} \quad (3.44)$$

where:
$\%(O_2)_{out}$ = The oxygen concentration in the effluent gas stream (% $^v/_v$)

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[3] In order to solve the oxygen gas mass balance, in addition to knowing the influent air flow rate and oxygen concentration in the vent gas, the carbon dioxide concentration in the vent gas or the respiration quotient Y_{CO2} (defined in Section 3.2.4) needs to be known. During this investigation from 112 paired O_2, CO_2 data, a Y_{CO2} value of 0.70 mol $(CO_2)_{gen}/\text{mol}(O_2)_{ut}$ was determined and this value is incorporated in Eq 3.44.

During phase I, the aerobic reactor was operated under eight distinct steady state periods, which differed from each other with respect to the main operating parameters (see Table 4.2). For each period, the rate of biological heat generation was calculated from the steady state heat balance as described. The results are given in Table 4.9. The average oxygen utilisation rate calculated for each period is contained in Table 4.3. The specific heat yield calculated for each period is contained in Table 4.10. Averaging the Y_h data yielded:

$$Y_h = 12.8$$

Equation 3.206 Average Value Determined for the Specific Heat Yield Y_h MJ/kg(O_2)

The average value determined for Y_h during phase I compares favourably with the Y_h = 13.0 MJ/kg(O_2) value obtained by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor. This confirms the accuracy of the heat and oxygen mass balances performed on the Athlone aerobic reactor, the results and calculations of which are discussed in detail in Chapter 4.

Further evidence of the accuracy of the heat and oxygen mass balances was provided during phase II when oxygenation was with both air and pure oxygen. Based on the adjustments made to the heat balance and oxygenation equations described in the Sections above, a Y_h value of 12.8 MJ/kg(O_2) was confirmed but now over a much wider *OUR* and H_b range (see Figure 7.5).

3.10.4 The Direct Determination of the Biological Heating Rate

Once the specific heat yield Y_h is known then for design the rate of biological heat generation H_b can be calculated directly by estimating the oxygen utilisation rate *OUR* (or oxygen transfer rate *OTR* of the aeration device) and applying Eq 3.201. When oxygenation is with air alone, the oxygen utilisation rate (given by Eq 3.44) and consequently the biological heating rate can be determined by measuring the oxygen concentration in the effluent gas stream ($\%(O_2)_{out}$). The expression for the biological heating rate is as follows:

$$H_b = 12.8 \times 1.205 \times Q(AIR)_{in} \cdot \frac{32}{29} \cdot \frac{(21.0 - \%(O_2))}{(100 - 0.3 \times \%(O_2)_{out})}$$

Equation 3.207 The Rate of Biological Heat Generation H_b (MJ/h) whilst Oxygenation is with Air Only and accepting Y_{CO2}=0.70mol/mol.

When oxygenation is with air and pure oxygen, the oxygen utilisation rate OUR^{AIR+O2} (now given by Eq 3.91) and consequently the biological heating rate H_b^{AIR+O2} can be determined by measuring the oxygen concentration in the combined effluent gas stream $(\%(O_2)_{out}^{AIR+O2})$. The expression for the biological heating rate is as follows:

$$H_b^{AIR+O2} = 12.8M(O_2)_{in}^{O2} + 3.57Q(AIR)_{in} - \frac{\%(O_2)_{out}^{AIR+O2}\left(16.0Q(AIR)_{in} + 8.96M(O_2)_{in}^{O2}\right)}{100 - 0.3\%(O_2)_{out}^{AIR+O2}}$$

Equation 3.208 **The Rate of Biological Heat Generation H_b^{AIR+O2} (MJ/h) whilst Oxygenation is with Air and Pure Oxygen and accepting Y_{CO2} = 0.70mol/mol.**

3.10.5 Modelling Aerobic Reactor Performance

With Equations Eq 3.207 and Eq 3.208 each term in the steady state heat balance, including the biological heating rate, can now be calculated directly from the measurement of the operating parameters listed in Section 3.10.2 above. This enables the performance of the aerobic reactor to be modelled. For example, it is possible to predict the minimum retention time required to operate at 50°C as a function of ambient air temperature when oxygenation is with air only. Pure oxygen supplementation rates can be predicted which enable the reactor retention time to be reduced. These are just two examples of a number of situations which can be modelled in this way. Important operating parameters, such as retention time and reactor temperature, can be calculated from the input of data from the other operating parameters. Because of the importance of this aspect in evaluating the performance of the aerobic reactor, Chapter 5 deals specifically with modelling the performance of the aerobic reactor whilst oxygenation is with air alone. Chapter 8 deals with modelling aerobic reactor performance, for oxygenation with air and/or pure oxygen. Further, the model presented in Chapter 8 is expanded to incorporate, anaerobic digester heating requirements, volatile solids destruction, digester biogas production and final sludge stability.

A summary view of all the heat loss and gain terms for the aerobic reactor, for oxygenation with air alone and pure oxygen alone, is presented graphically in Figure 3.12 below. The aerobic reactor temperature is fixed at 50°C with an ambient temperature of 20°C.

Using air alone for oxygenation, the minimum retention time at which the reactor can operate is limited by the maximum oxygen transfer rate which the aeration device can deliver. Further restriction is placed on the process, by the high vapour heat loss rate H_v, which accounts for approximately 40% of the total heat loss.

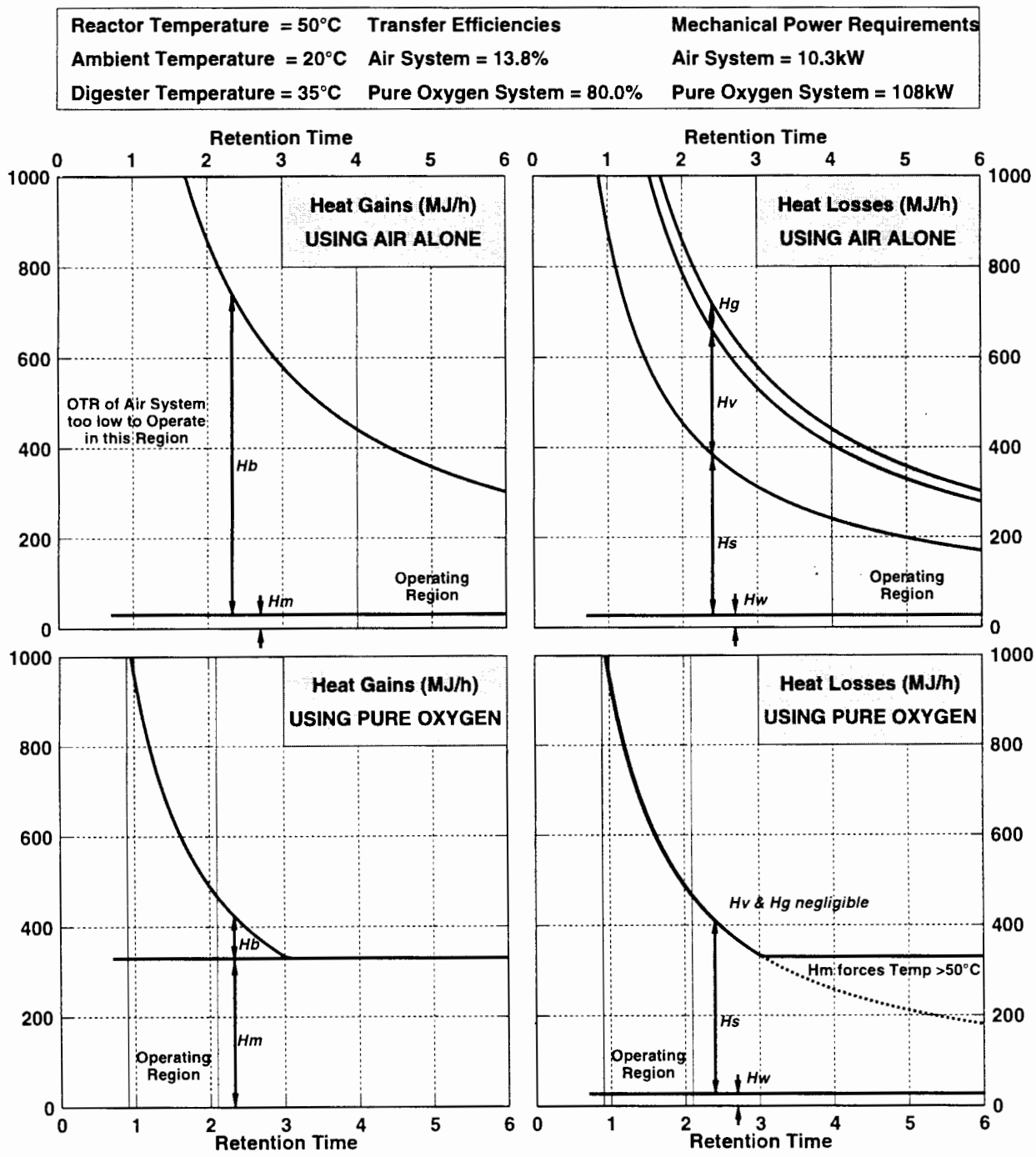

Figure 3.12 **Graphical Summary of the Heat Loss and Gain Terms for the Steady State Heat Balance across the Aerobic Reactor for Oxygenation with Air alone and Pure Oxygen alone. Variation with Retention Time.**

Using pure oxygen, the vapour heat loss rate is negligible. The high oxygen transfer rate which can be effected by the pure oxygen system makes operation at retention times as low as one day possible. The mechanical power requirement for pure oxygen injection is high. Indeed mechanical heating rate H_m is sufficient to support the process at retention times above 3 days.

3.11 VOLATILE SOLIDS DESTRUCTION

3.11.1 Introduction

For the purposes of this investigation the oxygen utilisation rate ***OUR*** was selected as the most appropriate parameter to enable the biological heating rate H_b to be determined. Consequently much of the discussion thus far in this Section has centred on establishing procedures for; (1) accurately measuring ***OUR***, and for (2) estimating the heat loss and gain terms in the steady state heat balance to allow H_b to be calculated by difference. There is however one other fundamental parameter, which is linked to the biological activity which takes place in the aerobic reactor and that is the **volatile solids destruction rate** $M(VS)_{dest}$

It is important for the dual digestion process, that the aerobic reactor is operated under **oxygen limiting conditions** (where the ***OUR*** is fixed by the maximum oxygen transfer rate OTR_{max} effected by the aeration device) so that the biological heat generation rate H_b is at its maximum. If insufficient substrate becomes available to the thermophilic organisms for them to sustain the biological heating rate at its maximum, either through a reduction in the feed sludge solids concentration or a lengthening of the sludge retention time, then the reactor is said to have become substrate limited. The ***OUR*** falls below OTR_{max} and positive D.O. levels will be recorded in the sludge. The need to be able to predict the point at which the reactor switches from oxygen limitation to substrate limitation, either in terms of feed sludge solids concentration or sludge retention time, is therefore important. In order to be able to predict the onset of substrate limitation, kinetic equations to describe VS destruction under both oxygen and substrate limiting conditions are developed. At the point at which the aerobic reactor becomes substrate limited both sets of kinetic equations will be valid. By equating these kinetic equations either the minimum feed sludge solids concentration or the maximum sludge retention time, to ensure that the reactor stays operating under oxygen limiting conditions, can be predicted.

The development of the relevent kinetic equations for identifying the transition from oxygen to substrate limitation is presented in this Section. Because ammonium production and alkalinity generation are linked to VS destruction, the Section concludes by estimating the increase in ammonia and alkalinity concentrations in the sludge liquid as a result of the breakdown of volatile solids in the reactor. The application of the formulae derived in this Section are presented in Chapters 5 and 8 on modelling of the aerobic reactor process.

3.11.2 The Relationship between VS Destruction and OUR

Accepting that the rate of volatile solids destruction in the aerobic reactor is proportional to the oxygen utilisation rate, the relationship between the two parameters can be described as follows:

$$M(VS)_{dest} = \frac{V_p OUR}{f_{ovs}} \quad \text{...kg(VS)/h} \quad (3.209)$$

where:

$M(VS)_{dest}$ = Rate of volatile solids destruction in the aerobic reactor (kg(VS)/h)
V_p = Aerobic reactor process volume (m^3)
OUR = Oxygen utilisation rate ($kg(O_2)/m^3.h$)
f_{ovs} = Mass of oxygen utilised per mass of VS destroyed ($kg(O_2)/kg(VS)$)

As discussed at the beginning of this Chapter (Section 3.1.3) the calculation of the **OUR** for ATAD systems is based on the proportionality between the ***OUR*** and $M(VS)_{dest}$. It was noted also that acceptance of this proportionality is reasonable for the ATAD process because the retention times are long (>5d) but that for the aerobic reactor in dual digestion, it may not be applicable because the retention times can be shorter. For the pure oxygen aerobic reactor the retention times is around 1 to 1.5 days and at retention times this short, Messenger *et al* (1992) found no proportionality between ***OUR*** and $M(VS)_{dest}$. However for the air oxygenated aerobic reactor, where retention times are longer (3 to 6 days), it is possible that the proportionality applies as it does for the ATAD system, and consequently this approach was evaluated in this investigation of the Athlone aerobic reactor. The theoretical development follows below.

3.11.3 Volatile Solids Destruction Under Oxygen Limiting Conditions

Under oxygen limiting conditions (substrate excess), the oxygen utilisation rate ***OUR*** is fixed by the maximum oxygen transfer rate OTR_{max} which the aeration device can deliver. It therefore follows that the rate of volatile solids destruction under oxygen limiting conditions is given by:

$$M(VS)_{dest} = \frac{V_p . OUR}{f_{ovs}} = \frac{V_p . OTR_{max}}{f_{ovs}}$$

Equation 3.210 The Rate of Volatile Solids Destruction in the Aerobic Reactor Under Oxygen Limiting Conditions $M(VS)_{dest}$ (kg(VS)/h)

For a specific OTR_{max}, both the rate of volatile solids destruction $M(VS)_{dest}$ and the oxygen utilisation rate ***OUR*** remain constant (independent of retention time and feed solids concentration) and the biological heating rate H_b is at its maximum. This aspect is illustrated in Figure 3.13 below, where the ***OUR*** under oxygen limiting conditions is indicated by a horizontal line at OTR_{max} (left hand side of Figure 3.13). The (reducing) effect on the ***OUR*** and $M(VS)_{dest}$ when the reactor becomes substrate limited (point A in Figure 3.13) is discussed in Section 3.11.4 below.

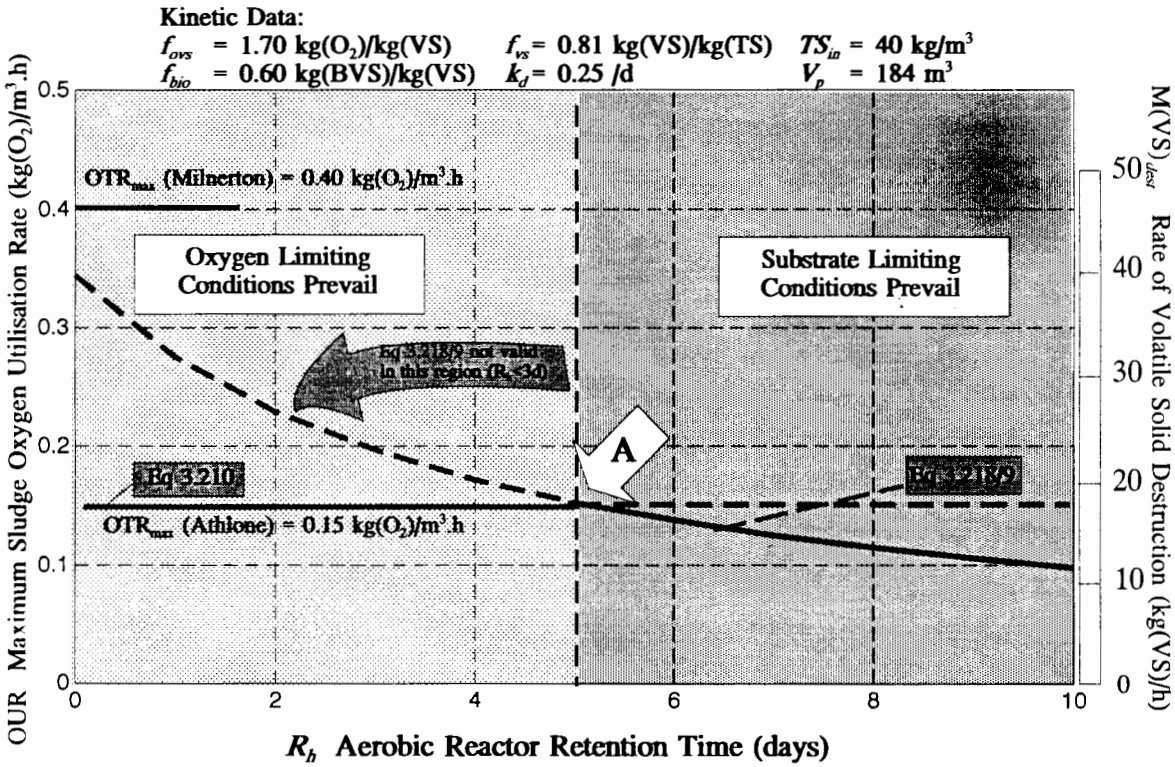

Figure 3.13 Maximum Sludge Oxygen Utilisation Rate and Volatile Solids Destruction Rate as a function of Retention Time under both Oxygen and Substrate Limiting Conditions.

The fraction of volatile solids removed in the aerobic reactor $f(VS)_{rem}^{O2lim}$ is defined as the ratio between the rate of volatile solids destruction $M(VS)_{dest}$ and the rate at which volatile solids enter the reactor $M(VS)_{dest}$ viz:

$$f(VS)_{rem}^{O_2\lim} = \frac{M(VS)_{dest}}{M(VS)_{in}} \quad \ldots \quad (3.211)$$

The rate of volatile solids entering the reactor is given by:

$$M(VS)_{in} = \frac{VS_{in}.Q(SL)_{in}}{24} \quad \text{...kg(VS)/h} \quad (3.212)$$

where:

VS_{in} = Concentration of volatile solids in the influent sludge (kgVS/m^3)

$Q(SL)_{in}$ = Influent sludge flow rate (m^3/d)

For modelling purposes it is more convenient to express; (1) the sludge flow in terms of the aerobic reactor process volume V_p and retention time R_h, and (2) the influent volatile solids concentration as a fraction f_{vs} of the influent total solids concentration viz.

$$Q(SL)_{in} = \frac{V_p}{R_h} \qquad \textit{and} \qquad VS_{in} = f_{vs}.TS_{in}$$

Equation 3.212 then becomes:

$$M(VS)_{in} = \frac{V_p.f_{vs}.TS_{in}}{24.R_h} \quad \text{...kg(VS)/h} \quad (3.213)$$

Substitution of Eq 3.213 for $M(VS)_{in}$ and Eq 3.210 for $M(VS)_{dest}$ into Eq 3.211 for $f(VS)_{rem}^{O2lim}$ yields:

$$f(VS)_{rem}^{O_2\lim} = \frac{24.R_h.OUR}{f_{ovs}.f_{vs}.TS_{in}}$$

Equation 3.214 The Fraction of Volatile Solids Removed in the Aerobic Reactor Under Oxygen Limiting Conditions $f(VS)_{rem}^{O2lim}$ (-)

This equation is important because it can be used to give an estimation of the likely reduction in volatile solids in the aerobic reactor, by measuring the oxygen utilisation rate and the influent sludge feed solids concentration. Under oxygen limiting conditions, the fraction of volatile solids removed in the aerobic reactor $f(VS)_{rem}^{O2lim}$ is proportional to the retention time (see Figure 3.14; left hand side) at a constant feed solids concentration, assuming ***OUR*** is fixed by OTR_{max}. The effect on the VS removal efficiency when the reactor becomes substrate limited (point A) is discussed below (Section 3.11.4).

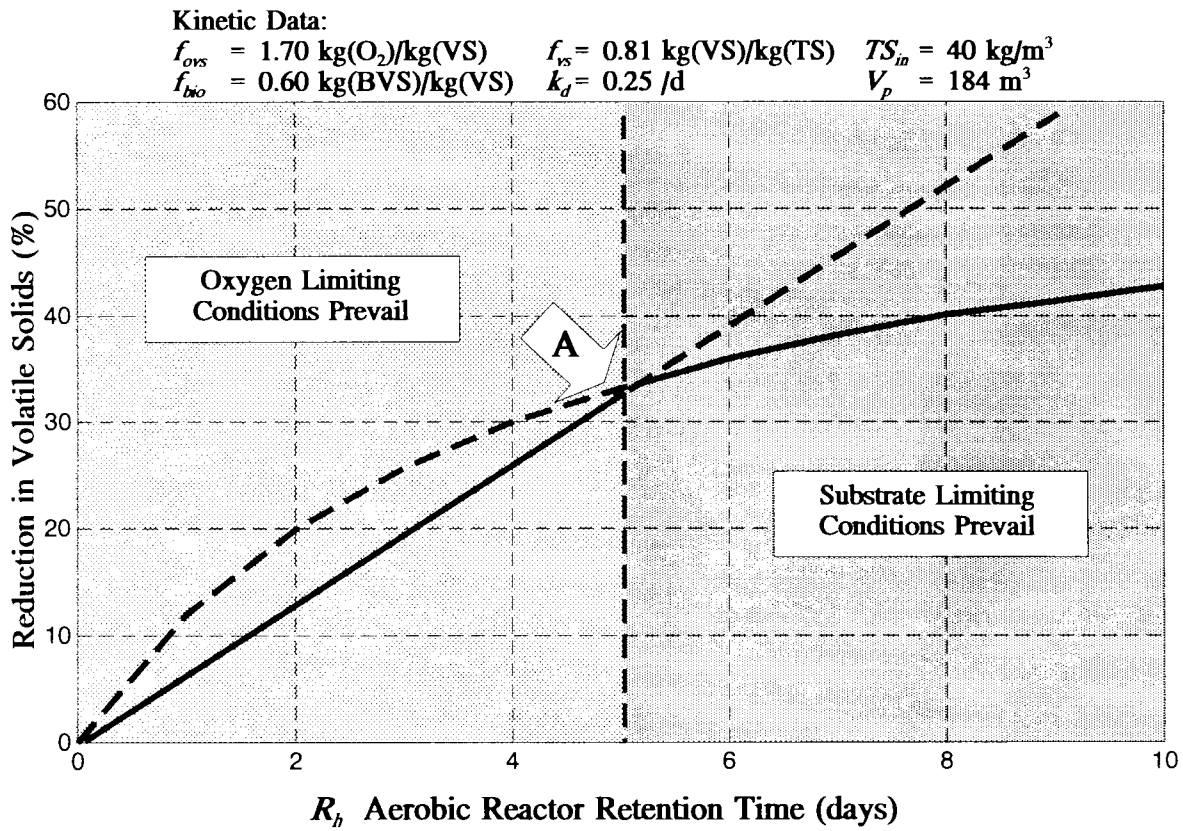

Figure 3.14 The Fraction of Volatile Solids Removed in the Aerobic Reactor as a function of Retention Time under both Oxygen and Substrate Limiting Conditions

3.11.4 Volatile Solids Destruction under Substrate Limiting Conditions

Under substrate limiting conditions (oxygen excess), the rate of volatile solids destruction was modelled by Andrews and Kambhu (1971) as a first order rate equation with respect to the biodegradable VS concentration (Eq 3.1) i.e.

$$M(VS)_{dest} = \frac{V_p}{24.R_h}([BVS]_{in} - [BVS]_{out}) = -\frac{k_d}{24}[BVS]_{out}V_p \quad \text{...kg(VS)/h} \quad (3.1)$$

where:

$M(VS)_{dest}$ = The rate of volatile solids destruction (kgVS/h)
k_d = The volatile solid destruction rate coefficient (/d)
R_h = The hydraulic retention time (d)
V_p = The effective process volume (m^3)
$[BVS]_{in}$ = The biodegradable VS concentration in the *influent* sludge (kgBVS/m^3)
$[BVS]_{out}$ = The biodegradable VS concentration in the *effluent* sludge (kgBVS/m^3)

It must be stressed that the kinetics of Andrews and Kambhu (1971) were modelled on the ATAD process, where the predominant bacterial activity is that of VS destruction. Whilst the model will be valid for reactor retention times in excess of 3 days[4], where VS destruction can be accepted to be the predominant activity, it cannot be used to predict substrate limitation at short retention times (<3d), where bacterial growth will become the more predominant factor. Under normal operating conditions for the air oxygenated aerobic reactor this would not be a problem (as retention times are longer than 3 days) and the kinetics of Andrews and Kambhu can be applied. However, if oxygen supplementation is considered as an option for reducing retention time, any reduction in R_h to less than 3 days would place the aerobic reactor under operating conditions where the VS destruction kinetics of Andrews and Kambhu are no longer applicable. Rearranging Eq 3.1 in terms of $[BVS]_{in}$ yields:

$$[BVS]_{out} = \frac{[BVS]_{in}}{1 + R_h.k_d} \quad \text{...kg(BVS)/m}^3 \quad (3.216)$$

Substitution of Eq 3.215 into the right hand side of Eq 3.1 allows the rate of volatile solids destruction to be expressed in terms of the influent biodegradable volatile solids concentration viz.

$$M(VS)_{dest} = \frac{k_d[BVS]_{in}V_p}{24.(1 + R_h.k_d)} \quad \text{...kg(VS)/h} \quad (3.217)$$

The influent biodegradable volatile solids concentration is linked to the total solids concentration in the following manner:

$$[BVS]_{in} = f_{bio}.f_{vs}.TS_{in} \quad \text{...kg(BVS)/m}^3 \quad (3.218)$$

where:

f_{bio} = Biodegradable fraction of the volatile solids

f_{vs} = Volatile solids fraction of the total solids

The rate of volatile solids destruction under substrate limiting conditions is therefore:

$$M(VS)_{dest} = \frac{k_d.f_{bio}.f_{vs}.TS_{in}.V_p}{24.(1 + R_h.k_d)}$$

Equation 3.218 The Rate of Volatile Solids Destruction Under Substrate Limiting Conditions $M(VS)_{dest}$ (kgVS/h); valid only for reactor retention times longer than about 3 days

[4] The retention time for ATAD systems is normally in excess of 3 days.

The oxygen utilisation rate ***OUR*** is linked to the rate of volatile solids destruction $M(VS)_{dest}$ by the constant of proportionality f_{ovs} (refer Eq 3.209). The oxygen utilisation rate under substrate limiting conditions is therefore given by:

$$OUR = \frac{k_d . f_{bio} . f_{vs} . TS_{in} . f_{ovs}}{24.(1 + R_h . k_d)}$$

Equation 3.219 The Oxygen Utilisation Rate Under Substrate Limiting Conditions *OUR* ($kg(O_2)/m^3.h$); valid only for reactor retention times longer than about 3 days

Under substrate limiting conditions, the rate of volatile solids destruction $M(VS)_{dest}$ and oxygen utilisation rate ***OUR*** decrease with increasing retention time (see Figure 3.12; curved line) for a constant feed solids concentration. In contrast, the $M(VS)_{dest}$ and ***OUR*** remain constant under oxygen limiting conditions.

The fraction of volatile solids removed by the process is obtained by dividing Eq 3.218 (the mass rate of VS destruction) by Eq 3.213 (the mass rate of VS entering the reactor) viz.

$$f(VS)_{rem}^{VS\,lim} = \frac{k_d . f_{bio} . R_h}{(1 + R_h . k_d)}$$

Equation 3.220 The Fraction of Volatile Solids Removed Under Substrate Limiting Conditions $f(VS)_{rem}^{VSlim}$ (-); valid only for reactor retention times longer than about 3 days

For comparison purposes, the change in the fraction of volatile solids removed in the aerobic reactor, under both oxygen $f(VS)_{rem}^{O2lim}$ (straight line) and substrate $f(VS)_{rem}^{VSlim}$ (curved line) limiting conditions, with retention time is illustrated in Figure 3.14 above.

3.11.5 Estimation of the Retention Time at Which Substrate Limitation Occurs

It is essential for the aerobic reactor to be operated under oxygen limiting conditions in order for the oxygen utilisation rate, and therefore the biological heating rate to be at its maximum. If however, the feed sludge solids concentration decreases or the sludge retention time is lengthened so that the reactor becomes substrate limited, the oxygen utilisation rate and biological heating rate will decrease (see Figure 3.13). It is desirable therefore to be able to predict the point (point A in Figures 3.13 and 3.14) at which this transition will take place in order to safeguard against its occurrence.

At the transition stage between oxygen limiting conditions and substrate limiting conditions, the rate of volatile solids destruction described by each set of kinetics (substrate and oxygen limiting) must be equal. i.e.

Oxygen Limiting (Eq 3.210) **Substrate Limiting (Eq 3.218)**

$$M(VS)_{dest} = \frac{V_p.OUR}{f_{ovs}} = \frac{k_d.f_{bio}.f_{vs}.TS_{in}.V_p}{24(1 + R_h.k_d)} \quad \text{...kg(VS)/h} \quad (3.221)$$

Rearranging in terms of the sludge retention time $\mathbf{R_h}$ yields

$$R_{h\,max} = \frac{f_{ovs}.f_{bio}.f_{vs}.TS_{in}}{24.OUR} - \frac{1}{k_d}$$

Equation 3.222 The Maximum Sludge Retention Time at Which the Aerobic Reactor can Operate Before it Becomes Substrate Limited $R_{h\,max}$ (d); valid for R_h>3d

This equation predicts the retention time at which substrate limitation will start to take place, conversely the VS destruction rate balance can be expressed in terms of feed sludge concentration viz.

$$TS_{in_{min}} = \frac{24.OUR.(1 + R_h.k_d)}{k_d.f_{ovs}.f_{bio}.f_{vs}}$$

Equation 3.223 The Minimum Feed Sludge Total Solids Concentration Required to Prevent Substrate Limitation Taking Place $TS_{in\,min}$ (kg(TS)/m^3); valid for R_h>3d

This equation predicts the minimum feed sludge concentration required to ensure that substrate limitation does not take place, for a specific retention time and oxygen utilisation rate.

In summary, the rate of volatile solids destruction $M(VS)_{dest}$, for any set of operating conditions, will be equal to the lesser of that given by Eq 3.210 (for oxygen limiting conditions) or Eq 3.218 (for substrate limiting conditions).

3.11.6 Estimation of the Increase in the Ammonium Concentration

Ammonia is released into solution in the aerobic reactor, as a result of the breakdown of volatile solids which contain nitrogen (see Section 3.1.9). An estimation of the increase in the ammonium concentration as a result of this breakdown is based on the (ammonia) nitrogen mass balance across the aerobic reactor viz

$$M(NH_4^+\text{-}N)_{out} = M(NH_4^+\text{-}N)_{in} + M(NH_4^+\text{-}N)_{gen} \quad \text{...g(N/h)} \quad (3.224)$$

where:

$M(NH_4^+\text{-}N)_{in}$ = Mass rate of (ammonia) nitrogen entering the reactor in the influent sludge (g(N)/h)

$M(NH_4^+\text{-}N)_{out}$ = Mass rate of (ammonia) nitrogen leaving the reactor in the effluent sludge (g(N)/h)

$M(NH_4^+\text{-}N)_{gen}$ = Mass rate of (ammonia) nitrogen generated in the reactor through the breakdown of volatile solids (g(N)/h)

The mass balance can be rewritten in terms of sludge flow rate and concentration as follows: It is assumed that the sludge flow rate remains constant through the reactor (i.e. that the reduction in volume due to vaporisation is negligible)

$$[NH_4^+]_{out}.Q(SL)_{in} = [NH_4^+]_{in}.Q(SL)_{in} + 24.M(NH_4^+\text{-}N)_{ge} \quad \text{...g(N)/h} \quad (3.225)$$

where:

$[NH_4^+]_{in}$ = Concentration of (ammonia) nitrogen the influent sludge (g(N)/m^3)

$[NH_4^+]_{out}$ = Concentration of (ammonia) nitrogen the effluent sludge (g(N)/m^3)

$Q(SL)_{in}$ = Influent sludge flow rate to the aerobic reactor (m^3/d)

The mass rate of (ammonia) nitrogen generated is estimated by accepting that the volatile solids contain a fixed fraction of nitrogen, which is released when the volatile solids are destroyed viz.

$$M(NH_4^+\text{-}N)_{gen} = f_n.M(VS)_{dest} \quad \text{...g(N)/h} \quad (3.226)$$

where:

f_n = Fraction of nitrogen in the volatile solids (g(N)/kg(VS))

$M(VS)_{dest}$ = Rate of volatile solids destruction in the aerobic reactor (kg(VS)/h)

Under oxygen limiting conditions, the rate of volatile solids is described by Eq 3.210, which upon substitution into Eq 3.226 yields:

$$M(NH_4^+\text{-}N)_{gen} = \frac{f_n . V_p . OUR}{f_{ovs}} \quad \text{...g(N)/h} \quad (3.227)$$

Substitution of Eq 3.227 into Eq 3.224, rearranged to express the increase in (ammonia) nitrogen concentration yields:

$$[NH_4^+]_{out} - [NH_4^+]_{in} = \frac{24 . f_n . V_p . OUR}{f_{ovs} . Q(SL)_{in}} \quad \text{...g(N)/m}^3 \quad (3.228)$$

Expressing Eq 3.228 in terms of retention time yields:

$$[NH_4^+]_{out} - [NH_4^+]_{in} = \frac{24 . f_n . R_h . OUR}{f_{ovs}}$$

Equation 3.229 The Increase in Ammonia Nitrogen Concentration in the Sludge as a Result of the Destruction of Volatile Solids in the Aerobic Reactor under Oxygen Limiting Conditions (g(N)/m^3)

Under substrate limiting conditions, the rate of volatile solids is described by Eq 3.218, which upon substitution into Eq 3.226 yields:

$$M(NH_4^+\text{-}N)_{gen} = \frac{f_n . k_d . f_{bio} . f_{vs} . TS_{in} . V_p}{24 . (1 + R_h . k_d)} \quad \text{...g(N)/h} \quad (3.230)$$

Substitution of Eq 3.230 into Eq 3.224, rearranged to express the increase in (ammonia) nitrogen concentration yields:

$$[NH_4^+]_{out} - [NH_4^+]_{in} = \frac{f_n . k_d . f_{bio} . f_{vs} . TS_{in} . V_p}{24 . (1 + R_h . k_d) . Q(SL)_{in}} \quad \text{...g(N)/m}^3 \quad (3.231)$$

Expressing Eq 3.228 in terms of retention time yields:

$$[NH_4^+]_{out} - [NH_4^+]_{in} = \frac{f_n . k_d . f_{bio} . f_{vs} . TS_{in} . R_h}{24 . (1 + R_h . k_d)}$$

Equation 3.232 The Increase in Ammonia Nitrogen Concentration in the Sludge as a Result of the Destruction of Volatile Solids in the Aerobic Reactor under Substrate Limiting Conditions (g(N)/m^3)

CLOSURE

The equations developed above in Chapter 3 are used in the evaluation of the measured data for the Athlone aerobic reactor, both for phase I of the investigation, where oxygenation was with air alone, and for phase II (where applicable), where oxygenation was with air plus pure oxygen.

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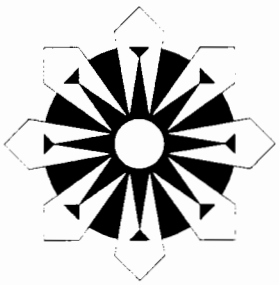

CHAPTER 4

RESULTS AND DISCUSSION: PHASE I

4.1 INTRODUCTION

The general aim of phase I of the research programme was to demonstrate the practicability of the dual digestion process, employing air rather than oxygen to successfully stabilise sewage sludge at short anaerobic digester retention times while achieving at the same time, sludge disinfection. The specific objectives of this phase of the investigation (see Section 1.4.1) were planned to cover all the claimed benefits of the dual digestion process. To satisfy these objectives the following aspects of the dual digester performance were evaluated in detail:

- the utilisation of oxygen within the aerobic reactor
- the aerobic reactor heat balance
- the biological heat generation rate within the aerobic reactor
- the effect of the foam layer depth on aerobic reactor performance
- the conditioning effects of aerobic pre-treatment
- the de-waterability of the final sludge
- the bacteriological quality of the final sludge
- the stability of the final sludge in terms of VS and COD reduction

Monitoring of the dual digester at Athlone for phase I commenced on the 5th October 1989 (designated day 1 of the evaluation period) and lasted for 45 weeks until the 12th August 1990 (day 312). During this period the dual digester operated continuously, without any major mechanical problems. A summary of the results obtained during phase I are presented and discussed in this Chapter. A complete compilation of the results is given in Appendix 2.

In order to present and discuss the details of the evaluation in a logical format, a summary of the overall plant performance is first given whereby the reader is made aware of the scale of plant operation prior to a detailed discussion on aerobic reactor performance. Thereafter a detailed discussion of aspects of the system performance is

presented. In order to obtain reliable data for the determination of the rate of oxygen utilisation and consequently the biological heat generation rate via the aerobic reactor heat balance, it was necessary to operate the dual digester for prolonged periods of time under differing sets of steady state conditions. The results measured during these steady state periods provided the necessary information to firstly satisfy the above objectives and secondly to calibrate mathematical models to simulate the performance of the aerobic reactor for design purposes (see Chapter 5).

4.2 SUMMARY OF OVERALL PLANT PERFORMANCE

4.2.1 Introduction

The average performance of the dual digestion plant during phase I, which includes all the data measured during the 312 day evaluation period, is summarised in Table 4.1. The variation in the main operating parameters during this period is shown in Figure 4.1. The shaded areas in Fig 4.1 depict the steady state periods of operation which are discussed in greater detail in Section 4.3 below.

Table 4.1 Summary of Overall Plant Performance During Phase: Oxygenation of the Aerobic Reactor is with Air Only.

Sludge Type		Feed Sludge		Aerobic Sludge		Anaerobic Sludge	
Parameter		mean	range	mean	range	mean	range
Temperature	°C	20	12-27	49	40-59	31	24-36
Total Solids	g/ℓ	45	24-88	35	15-55	24	6-70
Volatile Solids	g/ℓ	37	20-74	28	12-47	17	3-46
COD	g(O)/ℓ	64	21-99	43	15-79	26	7-72
pH	-	5.4	4.8-6.1	7.4	6.1-8.1	7.4	7.0-7.9

Process Stage		Aerobic Reactor		Anaerobic Digester	
Parameter		mean	range	mean	range
Sludge Retention Time	d	4.4	2.9-8.1	42	28-79
Air Flow Rate	m^3/h	690	200-1200	-	-

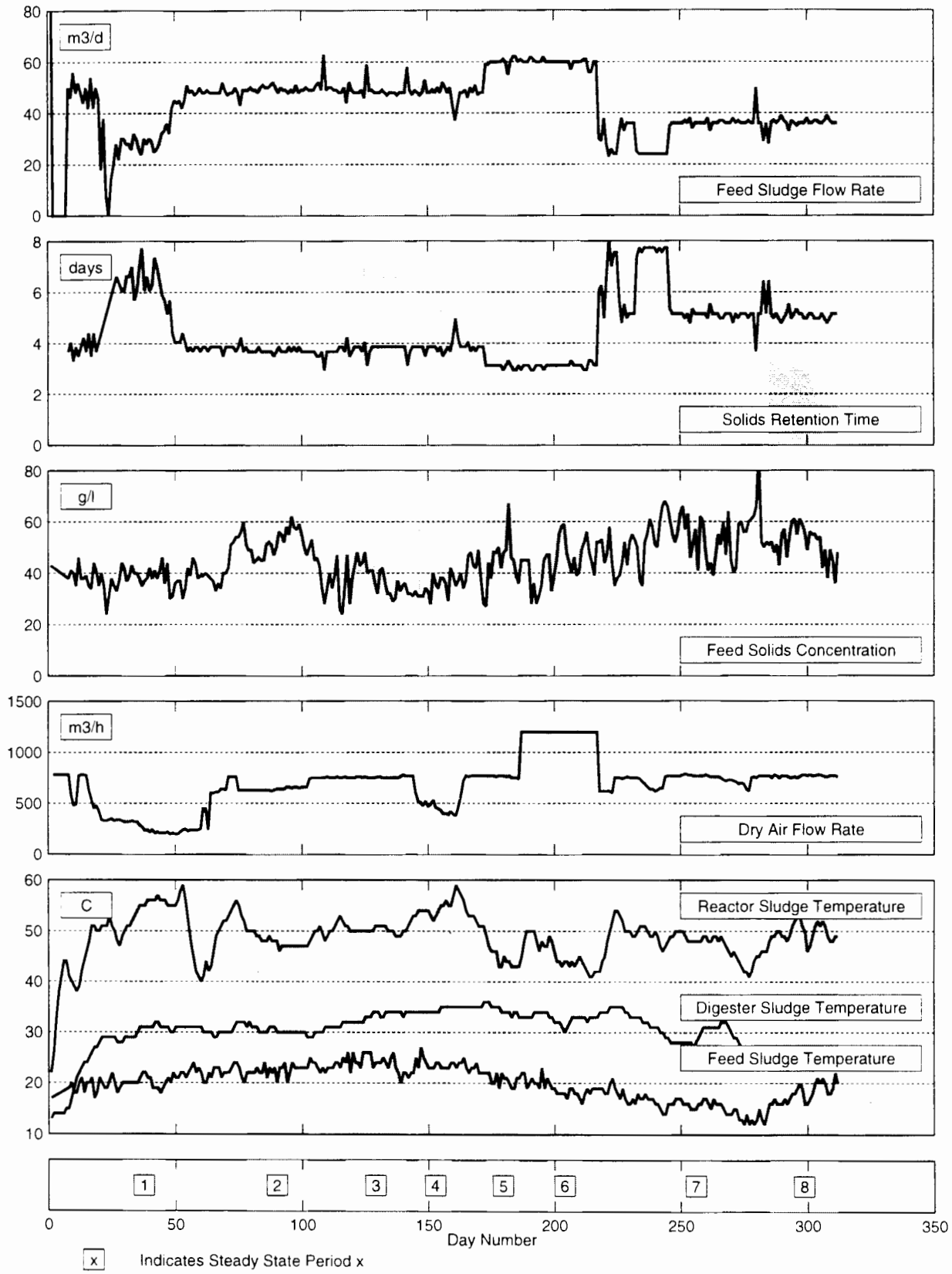

Figure 4.1 Graphical Representation of the Main Operating Parameters During Phase I: Oxygenation is with Air Only

A summary of the overall plant performance during phase I is discussed below, firstly in terms of the three major system operating parameters (process reliability, temperature and retention time), followed by a brief discussion on the effect of the process on the major chemical characteristics of the sludge.

4.2.2 Process Reliability

The process proved to be simple to operate and no major mechanical problems were encountered during the evaluation period. The electrical motors on the mixing pumps did, however, cause problems soon after the evaluation period ended. This aspect has been discussed in detail in Section 3.9 above.

Operating conditions in the aerobic stage were generally stable. The occasional drop in aerobic reactor temperature which took place was largely a result of either; (1) poorly thickened sludge (TS<30 kg/m^3) being fed to the reactor or; (2) excessive quantities of foam being spilled from the roof of the reactor. The effect, in both cases, was that the reactor became substrate limited; confirmed by the high (>2 $mg(O_2)/\ell$) dissolved oxygen levels recorded in the sludge at these times.

A sudden drop in reactor temperature occurred shortly after steady state period 1 ended (day 52), which could not be attributed to either of the above causes, but rather was suspected to be due to an incident of cadmium poisoning.[1]

4.2.3 Aerobic Reactor Operating Temperature

The average aerobic reactor temperature recorded over the full period of phase I was 49°C. The maximum and minimum recorded reactor temperatures during the period were 59°C and 40°C respectively (see Table 4.1). The variation observed in reactor temperature was largely due to the enforced changes made to the air flow rate and sludge flow rate in order to achieve the different steady state conditions (see Figure 4.1). This variation is not indicative of process instability, but rather a consequence of; (1) the changes made to the specific process parameters for research purposes; and (2) seasonal effects due to the temperature change in the feed sludge. The effect of the different operating conditions on reactor temperature is governed by the steady state heat balance.

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[1] A high level of cadmium was detected in the raw sewage composite sample on the day in which the drop in aerobic temperature occurred. With all operating parameters being normal it was suspected that the drop in temperature was due to the thermophilic bacteria being poisoned by the incoming cadmium.

4.2.4 Anaerobic Digester Operating Temperature

The average anaerobic digester temperature recorded over the full period of phase I was 31°C. Because the only heat source for the digester was the hot reactor sludge, the digester temperature was generally below the desired mesophilic value of 37°C and varied approximately seasonally with 36°C in summer and 24°C in winter (Table 4.1).

4.2.5 Aerobic Reactor Retention Time

The average aerobic reactor retention time over the full period of phase I was 4.4 days and varied between 2.9 and 8.1 days (Table 4.1). The variations in reactor retention time (as with reactor temperature) were variations; (1) imposed on the reactor as part of the research requirements and; (2) due to seasonal temperature changes in the feed sludge and ambient air (Figure 4.1). They were not a consequence of inherent system instability.

4.2.6 Anaerobic Digester Retention Time

The average anaerobic digester retention time over the full period of phase I was 42 days and varied between 28 days and 79 days (Table 4.1). The retention times which were achieved in the digester were a consequence of the sludge feed volume, selected to achieve the desired retention time in the aerobic reactor (Table 4.1). Because of the relatively large operating volume of the digester (1800m^3) it was not possible to reduce the retention time of the digester below 12 days to test the claim made for the process that **aerobic pre-treatment reduces the minimum required anaerobic digester retention time to achieve sludge stabilisation**. Consequently the claim was tested at laboratory scale (for results see Section 4.5 below). During phase II of the investigation, when the the oxygenation system was supplemented with pure oxygen, much higher loading rates were possible ($R_h \approx 1day$), consequently the retention time in the anaerobic digester could be lowered to below 10 days. In addition, due to the high sensible heat content of the hot aerobic sludge passing to the anaerobic digester, the digester operated at thermophilic temperatures. A full discussion on dual digester performance during this period is presented in Chapter 7.

4.2.7 Change in Sludge Characteristics during Dual Digestion

On average, over the full period of phase I, approximately 25% of the volatile solids in the feed sludge were destroyed by treatment in the aerobic reactor, with a further 40% reduction by anaerobic digestion. The average overall percentage reduction in volatile solids across the process was 56% (Table 4.1). A full discussion on the system total solids, volatile solids, and COD reduction is given in Section 4.4.1 below.

The conditioning effects of aerobic treatment was noticeable by; (1) an increase in the average primary sludge feed pH of 5.4 to 7.4 after aerobic treatment; and (2) an increase

in the average ammonia and bicarbonate alkalinity levels from 113 mg(N)/l and 40 mg($CaCO_3$)/l to 365 mg(N)/l and 820 mg($CaCO_3$) respectively after aerobic treatment. Aspects of aerobic conditioning are discussed in detail in Section 4.4.3 below.

4.3 EVALUATION OF AEROBIC REACTOR PERFORMANCE

4.3.1 The Steady State Periods

For the purposes of the heat balance calculations over the aerobic reactor, phase I was divided into eight distinct steady state periods of stable operation which differed from each other with respect to:

- air flow rate to the aerobic reactor;
- feed sludge temperature and flow rate to the aerobic reactor;
- ambient temperature.

Table 4.2 Average Values of the Aerobic Reactor Operating Parameters for the Eight Steady State Periods of Operation During Phase I

No:	Period Dates	Period Days Nos	Sludge flow m^3/d	Aerobic SRT days	Feed Conc g/ℓ	Air Flow m^3/h	Feed sludge temp °C	Aerobic sludge temp °C
1	89-11-02 to 89-11-19	29-46	29	6.5	39	230	20	56
2	89-12-21 to 90-01-14	78-102	49	3.8	56	660	23	47
3	90-01-29 to 90-02-22	117-141	48	3.8	34	760	23	50
4	90-02-27 to 90-03-12	146-159	49	3.7	35	440	23	54
5	90-03-26 to 90-04-08	173-186	60	3.1	47	760	21	44
6	90-04-19 to 90-05-03	197-211	60	3.1	48	1200	18	44
7	90-06-07 to 90-06-27	246-266	36	5.1	49	760	16	48
8	90-07-16 to 90-08-12	285-312	37	5.0	47	770	20	49

The difference in operating parameters between steady state periods are listed in Table 4.2 and can be seen graphically in Figure 4.1. During each of the eight periods all the controllable operating parameters were held as constant as practically possible to allow steady state conditions to develop. The time gaps between each of the eight steady state periods allowed the dual digester to reach a new steady state after a change in one or more of the operating parameters. On some occasions it was necessary to adjust the

process operating conditions a number of times before a new steady state period commenced. e.g. if the airflow rate was increased too much, severe foaming sometimes occurred which could only be reduced by decreasing the air flow rate to some lower value (see Section 4.2.4). The average data for each steady state period were calculated on the last 10 days data of each particular period to allow establishment of steady state conditions. (This applies to the data in Table 4.2 also.)

4.3.2 Measured Steady State Aeration Characteristics

Expressions for the supply, transfer, and utilisation of oxygen in the aerobic reactor were derived (see Section 3.2) in terms of the measurable parameters i.e.

$Q(AIR)_{in}$ = The volumetric air flow rate into the aerobic reactor (m^3(STP)/h)
$\%(O_2)_{out}$ = The oxygen concentration in the effluent vent gas (% $_v$/^{v})
$\%(CO_2)_{out}$ = The carbon dioxide concentration in the effluent vent gas (% $_v$/^{v})

By measuring the carbon dioxide fraction in the vent gas, it was possible to determine the respiration quotient Y_{CO2}, defined as the number of moles of carbon dioxide generated per mole of oxygen utilised (see Section 2.4.9). Because of the minimal spread in the measured Y_{CO2} values (see Section 4.3.6), a constant value could be accepted for Y_{CO2} which allowed the $\%(CO_2)_{out}$ term to be eliminated from the equations for ***OUR*** and ***OTE*** (see Section 3.2.7). The equations derived in Section 3.2 are as follows:

$$\boldsymbol{OSR} = 0.00151 \times \boldsymbol{Q}(AIR)_{in} \quad \text{...kg}(O_2)/m^3.h \quad (3.47)$$

$$\boldsymbol{OUR} = \frac{\boldsymbol{Q}(AIR)_{in}}{138} \cdot \frac{(21 - \%(O_2)_{out})}{(100 - 0.3 \times \%(O_2)_{out})} \quad \text{...kg}(O_2)/m^3.h \quad (3.44)$$

$$\boldsymbol{OTE} = \frac{(100 - 4.76 \times \%(O_2)_{out})}{(100 - 0.3 \times \%(O_2)_{out})} \times 100 \quad \text{...\%} \quad (3.49)$$

The average oxygen supply rate ***OSR***, oxygen utilisation rate ***OUR*** and oxygen transfer efficiency ***OTE*** calculated for each of the steady state periods are shown in Table 4.3. A graphical representation of the variation in the aeration parameters throughout phase I is given in Figure 4.2.

Table 4.3 **The Average Air Flow Rate, *OSR*, *OUR*, and *OTE* for Each Steady State Period Operated during Phase I: Oxygenation with Air Only**

No	Air Flow Rate m^3/h	*OSR* $kg(O_2)/m^3.h$	*OUR* $kg(O_2)/m^3.h$	*OTE* %
1	220	0.345	0.112	32.4
2	660	0.997	0.123	12.3
3	760	1.149	0.151	13.1
4	420	0.675	0.167	24.7
5	760	1.154	0.143	12.4
6	1200	1.824	0.181	9.9
7	760	1.163	0.145	12.5
8	770	1.169	0.149	12.8

During each of the steady state periods (but not necessarily on occasions between the steady state periods, see Section 4.2.2 above) there was no limitation on available substrate, verified by the fact that the dissolved oxygen level in the sludge was consistently less than 0.1 mgO/l. Under such conditions, where sufficient substrate is available, *the reactor is said to be oxygen limited* and the rate at which oxygen is utilised by the bacteria (***OUR***) is fixed by the rate at which oxygen is transferred into solution by the aeration device (***OTR***), i.e. the biological oxygen utilisation rate is constrained by the oxygen transfer rate of the aeration system i.e. $OTR = OUR < OUR_{max}$

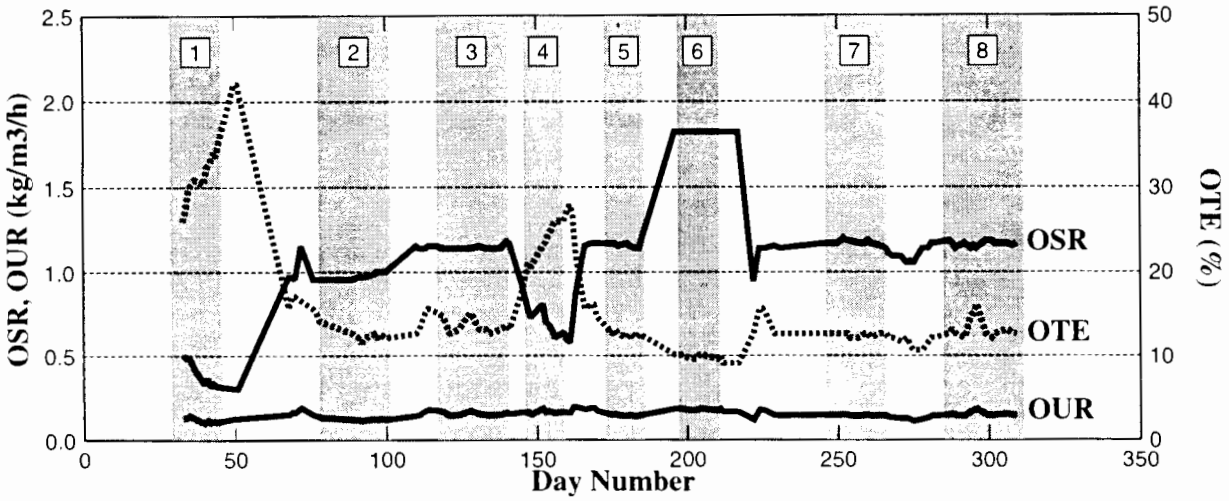

Figure 4.2 **The Variation in Oxygenation Characteristics During Phase I: Oxygenation with Air Only**

4.3.3 Aeration Characteristics during Non-Foaming Periods

During steady state periods 3,5,7, and 8, the influent air flow rate $Q(AIR)_{in}$ was set at approximately 760 m^3(STP)/h (the maximum air flow from one compressor) which yielded an ***OSR*** of around 1.16 kg(O_2)/m^3.h. No significant foaming occurred during these periods, which allowed the compressor to be operated at the maximum possible air flow rate. The measured values for ***OSR***, ***OUR***, and ***OTE*** showed little deviation between each of these periods (see Table 4.3) and the average values are summarised in Table 4.4.

Table 4.4 Measured Aerobic Reactor Aeration Characteristics *OSR*, *OUR*, and *OTE* During Periods (3,5,7,8) of Non-Foaming with an Air Flow Rate of 760 m^3(STP)/h

Aeration Characteristic		Average Value
Oxygen Supply Rate (***OSR***)	kg(O_2)/m^3.h	1.159
Oxygen Utilisation Rate (***OUR***)	kg(O_2)/m^3.h	0.147
Oxygen Transfer Efficiency (***OTE***)	%	12.7

The recorded average value of 0.147 kg(O_2)/m^3.h for ***OUR*** is approximately 10% below the calculated OTR_{max} of 0.167 kg(O_2)/m^3.h determined in the non-steady state aeration test (refer Appendix 4 or Section 2.3.4). The principal difference in conditions between the aeration test and that encountered in the evaluation are:

- Sludge instead of water is being aerated. The mass transfer coefficient K_{La} will be reduced by the presence of (1) impurities in the sludge e.g. surface active agents that can absorb at the air-liquid interface, and (2) the concentration of solids in the sludge, which increase the liquid viscosity.

- The higher temperatures encountered during the evaluation (50°C compared to 20°C). The rate of oxygen transfer is affected in various ways by an increase in temperature. The principal effects are; (1) a decrease in the saturation concentration of oxygen because of the reduced oxygen solubility in the water (sludge), (2) an increase in the mass transfer coefficient K_{La} due to the increased rate of diffusion of oxygen from the gas phase to the liquid phase, and (3) a reduction in the partial pressure of oxygen in the gas phase due to the increase in the humidity of the gas stream (at 50°C this is close to 10%).

Fuggle and Spensley (1985) state that the oxygen transfer capability of the aeration system may be reduced by 10% at 50°C compared to operation at 20°C. This is in agreement with the observations made on the aerobic reactor during this evaluation. For design purposes a value of 0.15 $kg(O_2)/m^3.h$ for ***OUR*** will be used in Chapter 5 for modelling the performance of the aerobic reactor under periods of no foaming with the air flow rate set at 760 $m^3(STP)/h$.

4.3.4 Aeration Characteristics with Two Compressors

During Steady State period 6, two compressors were employed to supply air to the aerobic reactor. This produced an average air flow rate of $1200m^3(STP)/h$. No foam was present on the surface of the reactor during this period. Table 4.5 compares the aeration characteristics observed during this period with the average of those measured at 760 $m^3(STP)/h$ with one compressor in operation (periods 3,5,7 and 8, Table 4.4), both under non-foaming conditions:

Table 4.5 Comparison Between Measured Aerobic Reactor Aeration Characteristics *OSR, OUR,* and *OTE* Obtained with One and Two Compressors in Operation

Aeration Characteristic		One Comp.	Two Comp.	Percent Change
Air Flow Rate ($Q(AIR)_{in}$)	$m^3(STP)/h$	760	1200	+58%
Oxygen Supply Rate (***OSR***)	$kg(O_2)/m^3.h$	1.159	1.824	+58%
Oxygen Utilisation Rate (***OUR***)	$kg(O_2)/m^3.h$	0.147	0.181	+20%
Oxygen Transfer Efficiency (***OTE***)	%	12.7	9.9	-20%

Increasing the air flow rate $Q(AIR)_{in}$ from $760m^3(STP)/h$ to $1200m^3(STP)/h$ increased the oxygen supply rate ***OSR*** by 58%. However the oxygen utilisation rate ***OUR*** increased by only 20% due to the reduction in the oxygen transfer efficiency ***OTE*** at the higher air flow rate. The effect of these differences on the heat balance and reactor temperature is discussed in Section 4.3.9 below.

4.3.5 The Effect of Foam on the Aeration Characteristics

During steady state periods 1 and 4 large quantities of foam were present on the surface of the sludge. In both cases the foam depth was in excess of 3m. To avoid spillage and to control the depth of foam, the air flow rate was reduced. Controlling the foam level in this manner is described as *positive foam management*. This approach makes use of the fact that there is a dynamic relationship between the creation of the foam by the aeration

device and the destruction of the foam by the antifoam properties in the incoming sludge. Foam is permitted to accumulate on the reactor surface and achieve some sort of equilibrium in terms of its production and collapse.

Reducing the air flow rate, to reduce the foaming, lowered the ***OSR*** significantly (see Table 4.6 below). However a concomitant decrease in the ***OUR*** was not observed due to an observed increase in ***OTE*** during periods of heavy foaming (steady state periods 1 and 4 in Figure 4.2). This effect is illustrated in Figure 4.3. Because the foam layer affects the ***OTE***, it also affects the heat balance of the aerobic reactor; which is discussed in Section 4.3.10 after presenting the heat balance results.

Table 4.6 Comparison Between the Measured Aerobic Reactor Aeration Characteristics *OSR*, *OUR*, and *OTE* During Non-Foaming and Foaming Periods

Aeration Characteristic		Period 3,5,7,8	Period 1	Percent change	Period 4	Percent Change
Foam		no	yes		yes	
Air Flow Rate $Q(AIR)_{in}$	m^3(STP)/h	760	220	-71%	420	-42%
Oxygen Supply Rate ***OSR***	kg(O_2)/m^3.h	1.159	0.345	-71%	0.675	-42%
Oxygen Utilisation Rate ***OUR***	kg(O_2)/m^3.h	0.147	0.112	-24%	0.167	+14%
Oxygen Transfer Efficiency ***OTE***	%	12.7	32.4	+155%	24.7	+94%

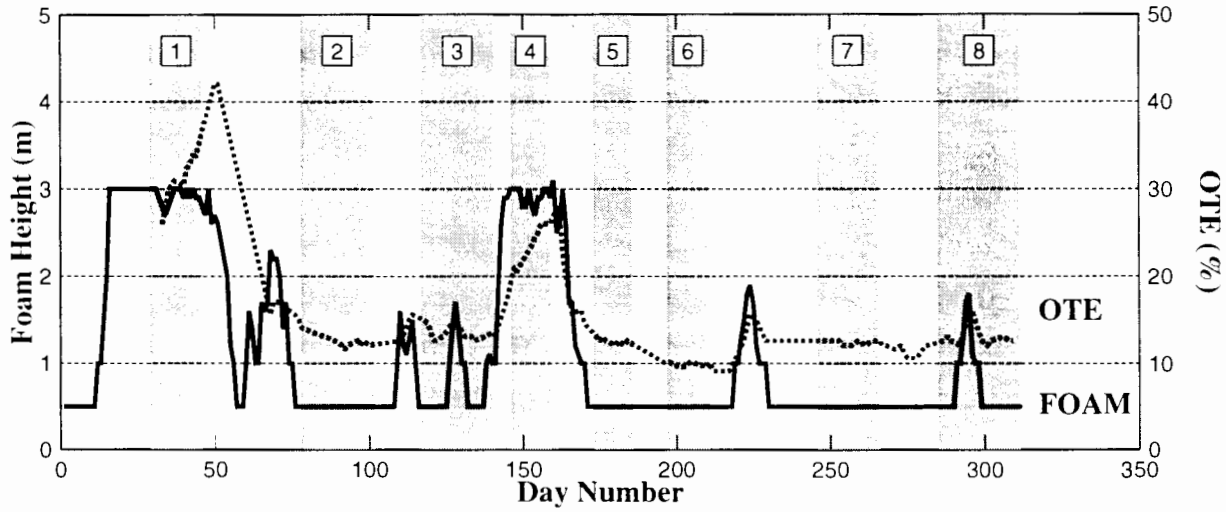

Figure 4.3 Effect of Foam Height on Oxygen Transfer Efficiency (OTE): Oxygenation is with Air Only

The aeration characteristics are considerably improved during periods of foaming. Nearly double the mass of oxygen supplied to the reactor can be transferred to the sludge liquid when foam is present in comparison to that which can be transferred when the foam is absent i.e. ***OTE*** almost doubles (see Table 4.6). If the foam layer could be positively managed so that the ***OSR*** does not have to be reduced to control it (as needed to be done at Athlone) then at the normal operating ***OSR*** of 1.16 $kg(O_2)/m^3.h$, with an ***OTE*** of 25%, the ***OTR*** would have been 0.29 $kg(O_2)/m^3.h$; a rate which is nearly double that measured in the non steady state aeration test (0.167 $kg(O_2)/m^3.h$. for a similar ***OSR***. It is appropriate therefore to consider why the aeration characteristics change under foaming conditions; in terms of the non-steady state aeration test, the mass transfer coefficient K_{La} would have to almost double to take account of foaming conditions.

The rate of oxygen transfer into solution is generally described by the two-film theory of Lewis and Whitman (1924) which, expressed in its more basic form, is as follows:

$$\frac{dC}{dt} = (D.S)^{1/2}.\left(\frac{A}{V}\right).(C_s - C) \quad \text{...mg}(O_2)/\text{l.s} \quad (4.1)$$

where:

D = diffusivity of oxygen into the liquid (m^2/h)
S = mean rate of exchange of the liquid film (surface renewal rate (h^{-1})
A = total interfacial area between the air phase and liquid phase (m^2)
C = dissolved oxygen concentration in the liquid ($mg(O_2)/l$)
C_s = saturation dissolved oxygen concentration in the liquid ($mg(O_2)/l$)

Under practical test conditions it is impossible to determine (amongst other things) the interfacial area A between the air and the liquid phases. It is usual, therefore, to use the overall coefficient K_{La} (which is measured in the non-steady state aeration test) where:

$$K_{La} = (D.S)^{1/2}.\left(\frac{A}{V}\right) \quad \text{...s}^{-1} \quad (4.2)$$

The effect of (1) solids concentration, (2) the impurities in the sludge and (3) the difference in temperature on the mass transfer coefficient K_{La} were all considered above in Section 4.3.3. The 10% difference between the observed OTR_{max} under non-foaming conditions during the evaluation and the OTR_{max} measured during the aeration test is in agreement with the claims made by Fuggle and Spensley (1985). However, the nonsteady state aeration test cannot distinguish (i.e. predict the effect on K_{La}) between foaming and non-foaming conditions.

Under foaming conditions, air bubbles can be considered to be entrained (trapped) in the sludge, thereby increasing the contact time between the liquid and gas phases and greatly increasing the interfacial area between the gas phase and the liquid phase. As both of these factors are incorporated into the K_{La} value, an increase in the K_{La} value can be expected under foaming conditions, which would then give rise to an increase in OTR_{max} which could then account for the higher *OTE*'s observed under foaming conditions.

It was not known under foaming conditions whether a defined liquid-foam interface remained at the 6m liquid level in the reactor with a 3m foam layer above it, or whether the entire 9m of sludge became a liquid/foam suspension. On the basis that the former situation existed, the fact that the *OTE* doubled implies that 3m of foam has the same K_{La} value as 6m of liquid, with the result that the K_{La} in the foam is twice that of the liquid. On the basis of the latter situation, 9m of liquid/foam suspension has a K_{La} double that of the liquid with the result that the K_{La} of the liquid/foam suspension is about 50% greater than that of the liquid only. In either situation the improvement with 3m of foam is equivalent to 6m of liquid depth. This suggests a possible aeration improvement scheme with pure oxygen enriched air to reduce the aerobic reactor retention time to 1.5 to 3 days; viz. air is utilised to create a foam layer and pure oxygen is injected to increase the *OTR* (see Chapter 5, Section 5.4).

One can conclude at this stage that the non-steady state aeration test makes a good prediction of the maximum oxygen transfer rate OTR_{max} under non-foaming conditions. However, because of the (anticipated) change in the mass transfer coefficient K_{La} when foaming occurs, the maximum oxygen transfer rate OTR_{max} measured from the non-steady state aeration test cannot be used to predict OTR_{max} under foaming conditions.

4.3.6 The Respiration Quotient

Measurement of both the oxygen and carbon dioxide fractions in the vent gas during phase I allowed the respiration quotient Y_{CO2}, defined as the number of moles of carbon dioxide produced per mole of oxygen utilised by the bacteria, to be determined (see Section 3.2):

$$Y_{CO_2} = \frac{\%(CO_2)_{out}(100-\%(O_2)_{in})}{\%(O_2)_{in}(100-\%(O_2)_{out}-\%(CO_2)_{out}) - \%(O_2)_{out}(100-\%(O_2)_{in})} \quad \text{...kmol/kmol} \qquad (3.27)$$

The average measured oxygen and carbon dioxide concentrations in the vent gas, and the calculated respiration quotient Y_{CO2} for each steady state period are presented in Table 4.7 below.

Table 4.7 The Average Measured Dry Vent Gas O_2 and CO_2 Concentrations (% v/v) and the Calculated Respiration Quotients Y_{CO2} for each Steady State Period During Phase I: Oxygenation with Air Only

No	$\%(O_2)_{out}$	$\%(CO_2)_{out}$	Y_{CO2}	$Q(AIR)_{in}$	Foam
1	14.5	4.8	0.69	220	yes
2	18.6	1.8	0.68	660	no
3	18.4	2.0	0.73	760	no
4	16.1	3.7	0.70	420	yes
5	18.5	1.8	0.68	760	no
6	19.0	1.4	0.69	1200	no
7	18.5	1.9	0.72	760	no
8	18.5	1.9	0.70	760	no

The average respiration quotient calculated over the full period of phase I was:

$$Y_{CO2} = 0.70 \text{ mol } (CO_2)_{gen}/\text{mol}(O_2)_{ut}$$

In total, 118 pairs of data were collected during phase I. The average value for Y_{CO2} of 0.70 compares very favourably with the 0.67 value observed by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor. The fraction of the carbon dioxide generated which becomes dissolved in the reactor sludge liquid, forming ammonium bicarbonate, is estimated to be less than 5% of the total mass generated. The effect of this reduction on the Y_{CO2} value is considered insignificant and therefore was ignored.

The distribution of Y_{CO2} values is shown in Figure 4.4. Because of the narrow spread in the observed Y_{CO2} values, a constant value of 0.70 was employed in the derivation of the relevant formulae to describe the oxygenation characteristics of the aeration system (see Section 3.2). Accepting a constant value for Y_{CO2} allowed elimination of the $\%(O_2)_{out}$ term in the final equations for ***OUR*** and ***OTE*** (see Equations 3.44 and 3.49).

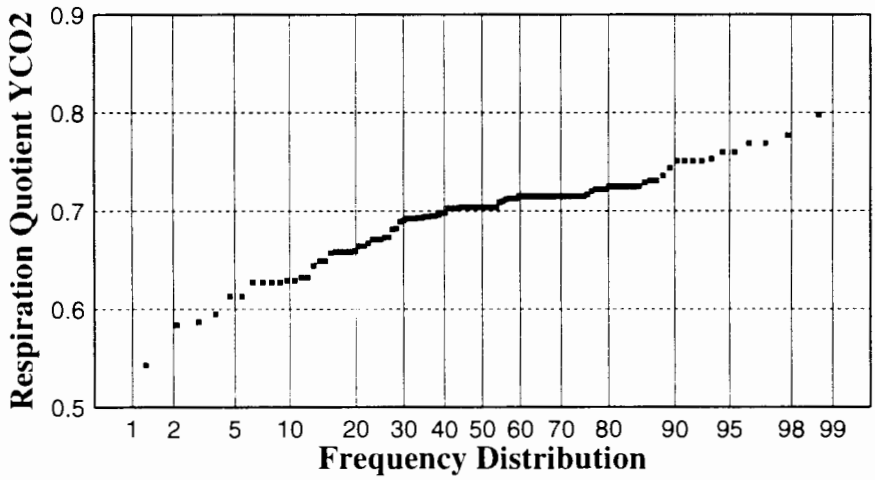

Figure 4.4 Frequency Distribution of Measured Respiration Quotient Values During Phase I: Oxygenation with Air

The Steady State Heat Balance

The heating performance of the aerobic reactor is assessed by establishing a steady state heat balance around the aerobic reactor. This is done by keeping the main operating parameters as constant as practically possible, thus allowing the reactor the reach steady state conditions, at which point the heat sources equal the heat sinks. The overall steady state heat balance (introduced in Section 3.1) is expressed as follows:

$$H_b + H_m = H_s + H_v + H_g + H_w \quad \text{...MJ/h} \quad (3.8)$$

where:

H_b = Rate of biological heat generation
H_m = Rate of heat energy input from the mixing device
H_s = Rate of sensible heat loss with the sludge leaving the reactor
H_v = Rate of vapour heat loss with the effluent gas stream
H_g = Rate of sensible heat loss with the effluent gas stream
H_w = Rate of heat loss through the walls of the reactor

With the exception of the biological heat term H_b, each term in the balance can be calculated from physical measurements made on the reactor. The biological heat is then calculated by difference with Eq (3.8) knowing the other heat loss and heat gain terms. Equations for the calculation of each of the heat balance terms were derived in Chapter 3 and are as follows:

The Water Vapour Heat Loss Rate in the Effluent Gas (Eq 3.76)

$$H_v = 2.38 \times 1.205\, Q(AIR)_{in} \left(\frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 - T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 - T(AIR)_{out}}\right)} - \frac{0.62 \times alog_{10}\left(9.12 - \frac{2307}{273 - T(AIR)_{in}}\right)}{1240 - alog_{10}\left(9.12 - \frac{2307}{273 - T(AIR)_{in}}\right)} \right)$$

The Effluent Gas Sensible Heat Loss Rate (Eq 3.94)

$$H_g = 1.21\,Q(AIR)_{in}\left(0.001\,(T(AIR)_{out} - T(AIR)_{in}) + \frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} 0.00187\,T(AIR)_{out}\right)$$

The Sludge Sensible Heat Loss Rate (Eq 3.102)

$$H_s = 4.04\left(\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{1.21\,T(SL)_r Q(AIR)_{in}}{1000} \times \frac{0.62 \times alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - alog_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{Q(SL)_{in} T(SL)_{in}}{24}\right)$$

The Rate of Heat Loss from the Walls of the Reactor (Eq 3.133)

$$H_w = 0.82\{(T(SL)_r - T(SL)_d\} + 0.49\{T(SL)_r - T(AIR)_{in}\}$$

The Rate of Mechanical Heat Input to the Reactor (Eq 3.140)

$$H_m = 1.65 \times I$$

All the above terms H_v, H_g, H_s, H_w and H_m have units of MJ/h. The symbols are defined in Table 2.5 above. In order to calculate each of the above heat terms, and hence solve the steady state heat balance, the average values (obtained from the last ten days of each steady state period) for the monitoring parameters are listed in Table 4.8 below.

Table 4.8 Average Steady State Values of the Monitoring Parameters Required for Solution of the Steady Stae Heat Balance During Phase I

No	Feed Sludge Flow $Q(SL)_{in}$	Feed Sludge Temp $T(SL)_{in}$	Reactor Sludge Tem $T(SL)_r$	Digester Sludge Temp $T(SL)d$	Influent Air Temp $T(AIR)_{in}$	Air Flow Rate $Q(AIR)_{in}$	Effluent Dry Air $\%O_2$ $\%(O_2)_{out}$	Effluent Dry Air $\%\ CO_2$ $\%(CO_2)_{out}$	Effluent Air Temp $T(AIR)_{out}$
	m^3/d	°C	°C	°C	°C	m^3(STP)/h	%(v/v	%(v/v)	°C
1	28.5	20.0	55.9	31.2	20.4	227	14.50	4.77	56.0
2	49.0	22.8	47.0	29.9	25.1	656	18.56	1.78	44.0
3	48.0	23.2	49.9	33.8	23.7	756	18.40	2.02	47.0
4	49.1	23.3	54.2	34.5	23.9	444	16.07	3.67	54.2
5	60.1	20.5	43.7	34.0	22.7	759	18.54	1.80	41.4
6	59.8	17.8	43.8	32.1	20.5	1200	19.04	1.44	41.0
7	36.0	16.2	48.5	30.8	15.4	765	18.53	1.88	45.8
8	36.7	19.9	50.0	26.0	15.6	769	18.47	1.90	47.0

From the data in Table 4.8 and the equations listed above, the heat balance terms for each steady state period were calculated and the results of those calculations are presented in Table 4.9. A computer programme was compiled to perform the required calculations and this programme is listed in Appendix 8.

Table 4.9 The Calculated Heat Balance Terms for Each Steady State Period

NO	Heat Output Terms				Heat Input Terms	
	H_s	H_v	H_g	H_w	H_m	H_b
1	163	74	13	38	31	257
2	188	94	19	25	31	295
3	200	135	27	26	31	357
4	240	126	22	31	31	388
5	224	94	21	18	31	325
6	246	149	35	21	31	420
7	181	137	33	31	31	351
8	170	147	35	37	31	358

4.3.7 Calculation of the Specific Heat Yield Coefficient

The biological heating rate is directly proportional to the oxygen transfer rate ***OTR***, which under oxygen limiting conditions, fixes the oxygen utilisation rate ***OUR*** (see Section 3.1). This relationship is expressed in equation form by the following:

$$H_b = Y_h.OTR.V_p = Y_h.OUR.V_p \quad \text{...MJ/h} \quad (3.4)$$

where:

Y_h = Specific heat yield in terms of oxygen utilisation (MJ/kg(O_2))
OTR = Oxygen transfer rate (Volume Specific) (kg(O_2)m^3.h)
OUR = Oxygen utilisation rate (Volume Specific) (kg(O_2)m^3.h)
V_p = Process volume (m^3)

The biological heating rate H_b was determined by difference from the steady state heat balance (Eq. 3.8) for each steady state period. The calculated values for H_b are given in the right hand column of Table 4.9 above. The oxygen utilisation rate ***OUR*** was determined for each steady state period, by performing an oxygen mass balance (see Section 3.2) across the aerobic reactor. The calculated values for ***OUR*** are given in Table 4.3. above. The linear relationship between the biological heating rate H_b and the oxygen

utilisation rate ***OUR*** is shown graphically in Fig. 4.5. The slope of the plot is (from Eq 3.4) equal to $Y_h \times V_p$. The calculated values for the specific heat yield Y_h for each steady state period are given in Table 4.10 below. Note that the aerobic reactor process volume V_p is equal to 184m^3.

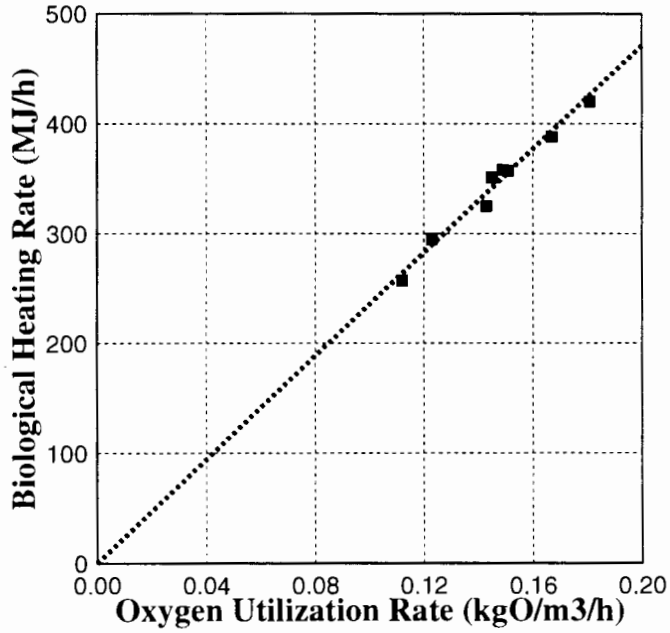

Figure 4.5 Biological Heating Rate versus Oxygen Utilisation Rate

Table 4.10 Calculated Specific Heat Yield Values for Each Steady State Period

No	*OUR* kg(O_2)/m^3.h	H_b MJ/h	Y_h MJ/Kg(O_2)
1	0.11	257	12.5
2	0.12	295	13.0
3	0.15	357	12.9
4	0.17	388	12.6
5	0.15	325	12.3
6	0.18	420	12.6
7	0.15	351	13.2
8	0.15	358	13.0

The calculated Y_h values vary in a very narrow band between 12.3 and 13.2 MJ/kg(O_2) with the average value being 12.8 MJ/kg(O_2). This average value compares favourably with the value of 13.0 MJ/kg(O_2) obtained by Messenger *et al* (1992), on a pure oxygen aerobic reactor, thereby confirming the accuracy of the tests performed at Athlone as

well as the assumptions in and generality of the mass and heat balance equations developed in Chapter 3.

If the scatter in the specific heat yield Y_h values had been large, then it could have been surmised that the heat balance equations which were employed for calculating Y_h, were not sufficiently general to account for the variation in operating conditions. The fact that the variation in Y_h values is small and that the average value obtained compares very favourably with Messenger *et al* (1992) provides confidence that not only are the equations sufficiently general, but that also they are reliably accurate.

4.3.8 The Effect of Operating with Two Compressors

Increasing the air flow rate to the aerobic reactor, by operating with two compressors, increased the oxygen supply rate ***OSR*** by 58% (see Section 4.3.4). As the effluent vent gas heat loss rates (H_v, H_g) are proportional to the ***OSR***, when the ***OSR*** was increased by 58% these heat losses, which make up about 40% of the total heat loss from the aerobic reactor (see Section 3.10.5; Figure 3.12), also increased by 58%. As the ***OUR*** only increased by 18%, the gain in the biological heating rate only increased by 18%. This heat gain is insufficient to compensate for the higher heat loss rates. Consequently the reactor temperature decreases until the heat losses are equal to the heat gain. This explains why it was not practical to operate with two compressors. Nevertheless, it was important to test this condition as it allowed the heat balance to be determined at a very high air flow rate, thereby broadening the range of operating conditions and validity of the heat and mass balance equations.

4.3.9 The Effect of Foam on Aerobic Reactor Heat Balance

In Section 4.3.5 the presence of a significant foam layer on the surface of the sludge in the aerobic reactor was shown to significantly increase the oxygen transfer efficiency *OTE* (see Figure 4.3). The effect of the foam layer on the aerobic reactor heat balance is illustrated in the following example: Comparing data for steady state periods 3 and 4 (Table 4.8), the ambient temperature and feed sludge temperature and flow rate are similar and accordingly one would expect the reactor temperatures to be similar also. However a significantly higher temperature is recorded during steady state period 4 (54.2°C) compared to period 3 (49.9°C). This difference is attributable to the presence of a significant foam layer (3.0m) on the sludge surface (see Table 4.11).

Table 4.11 Comparison of Operating Data Between Foaming and Non-Foaming Conditions During Phase I: Oxygenation with Air Only

No	Sludge Flow m^3/d	Feed Sludge temp°C	Foam Depth m	Air Rate m^3/h	*OSR* $kg(0_2)/m^3.h$	*OUR* $kg(0_2)/m^3.h$	*OTE* %	Aerobic Sludge temp °C
3	48	23	<1	760	1.149	0.151	13.1	50
4	49	23	3m	440	0.675	0.167	24.7	54

During steady state period 3 the reactor operated at 49.9°C with no significant foam layer present on the surface of the sludge. The switch to steady state period 4 was forced on the system by the onset of foaming. To control the foam layer and avoid foam spillage from the reactor, the air flow rate was reduced. This immediately reduced the effluent gas vapour and sensible heat loss rates (H_v, H_g) from the reactor. Because of the increase in the oxygen transfer efficiency ***OTE*** (13.1%→24.7%) as a result of foaming, the oxygen utilisation rate ***OUR*** actually showed a slight increase (0.15→0.17kg(O_2)/m^3.h), inspite of the decrease in the oxygen supply rate ***OSR*** (1.15→0.68kg(O_2)/m^3.h).

The increase in the oxygen utilisation rate ***OUR*** accordingly gave rise to an increase in the biological heating rate H_b. As a result of the increase in H_b and the reduction in H_v and H_g (due to the reduction in the air flow rate) the heat balance (Eq 3.8) was no longer in equilibrium. With the heat input into the system exceeding the heat losses at this point the temperature of the sludge in the reactor increased from 49.9°C to 54.2°C at which temperature a new steady state heat balance was again achieved. i.e. the temperature where the heat losses were again equal to the heat sources.

The effect of foaming on aerobic reactor performance is illustrated by Fig 4.6. The variation in reactor temperature across the full evaluation period is plotted, with the periods when significant foaming occurred indicated by shaded areas. From an examination of Fig 4.6 two conclusions can be drawn: (1) the onset of foaming took place only when the reactor temperature reached 50°C, and (2) noticeably higher reactor temperatures were achieved during periods of foaming; these higher temperatures can be attributed to the reduction in air flow rate (to prevent foam spillage) and the increase in oxygen transfer efficiency ***OTE*** (see Section 4.3.5 above). The effect of the foam layer on the reactor performance is evaluated in detail in Section 5.3.2.

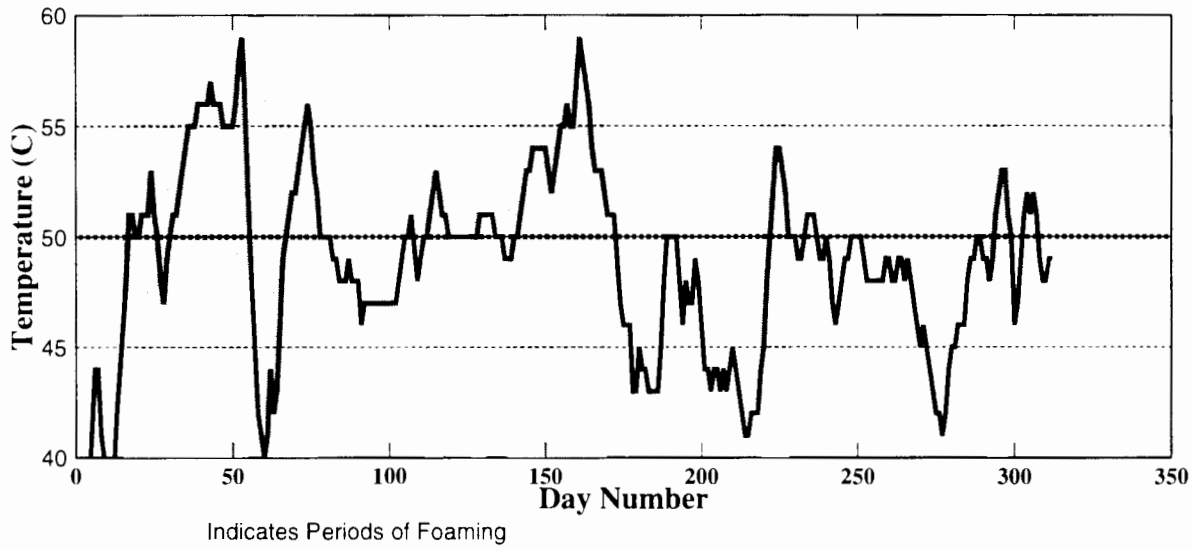

Figure 4.6 The Effect of Foaming on Aerobic Reactor Temperature

Whilst the presence of foam enhances the performance of the aerobic reactor, the problem of foam spillage causes practical operational problems. From this it can be concluded that without proper foam management and ***OSR*** control, operating temperatures in excess of 50°C are not practical. In this respect, the experience of foam management in ATAD systems will benefit the operation of the air oxygenated aerobic reactor to achieve and sustain temperatures greater than 50°C. However because ATAD systems invariably operate at 8 to 10 days retention time to achieve the required VS destruction (>38%), the ***OSR*** for these systems is considerably lower than in aerobic reactors of dual digesters, so foam control would in any event be less of a problem in ATAD systems than it is in air oxygenated aerobic reactors of dual digestion systems. An attempt to benefit from the advantages of a managed and controlled foam layer was made in Phase II of this investigation (see Section 6.3).

4.4 DETAILED ASPECTS OF OVERALL SYSTEM PERFORMANCE

4.4.1 System Volatile Solids and COD Reduction

Total Solids, Volatile Solids and COD analyses were performed on a daily basis throughout the 312 day evaluation period; with the exception of COD analysis on days 76-95. The average TS, VS, and COD concentrations in the feed, aerobic, and anaerobic sludges across the full evaluation period are given in Table 4.12 below. The objective of performing these analyses was to monitor the reduction of TS, VS, and COD across each stage of the system. It should be noted that in all the calculations below of TS, VS and COD removals, the loss of water from the reactor via vaporisation has not been taken into account; the assumption is made that the influent and effluent sludge flow rates are equal. This assumption is of course not valid because up to 4 to 7% of the water volume can be vaporised (depending on retention time) which effectively concentrates the aerobic reactor sludge. However, for the calculation of the VS and COD destroyed this effect is regarded to be small enough to be ignored considering the large variation in the daily VS and COD destroyed (removed) data; the latter being a consequence of the difference between two large and significantly variable measurements e.g. VS_{in} - VS_r (For the steady state heat balance, the effect of vaporisation was taken into account when calculating the sensible heat losses from the system - see Sections 3.5 and 3.6).

Table 4.12 Average TS, VS, and COD Concentrations in the Feed, Aerobic and Anaerobic Sludges Across the Full Evaluation Period: Phase I

Sludge Type	TS(g/ℓ)	VS(g/ℓ)	COD(g/ℓ)
Feed	45	36	64
Aerobic	34	27	43
Anaerobic	23	16	25

The degree of reduction is calculated over the full evaluation period (312 days). It was considered necessary to make the assessment over a prolonged period of operation in order to reduce the influence of; (1) the variation in solids concentration in the feed sludge from the gravity thickener; (2) the occasional foam spillages which occurred from the aerobic reactor; and (3) the variation in solids concentration in the final sludge from the anaerobic digester.

The variation in solids concentration in the final sludge from the anaerobic digester was a result of two effects: Firstly, the gas mixing in the anaerobic digester at the start of the

evaluation period was inefficient due to the low liquid operating level in the digester;[2] and secondly just prior to and during steady state period 6 (day 197 to 211) both liquid ring compressors were employed to supply air to the aerobic reactor, and therefore no gas mixing took place in the digester.

A graphical representation of the changes in TS, VS, and COD concentrations through each stage of the system is presented in Figure 4.7. The data have been smoothed statistically to eliminate short term fluctuations of less than 2 months by the use of a weighted moving average filter, prior to plotting.

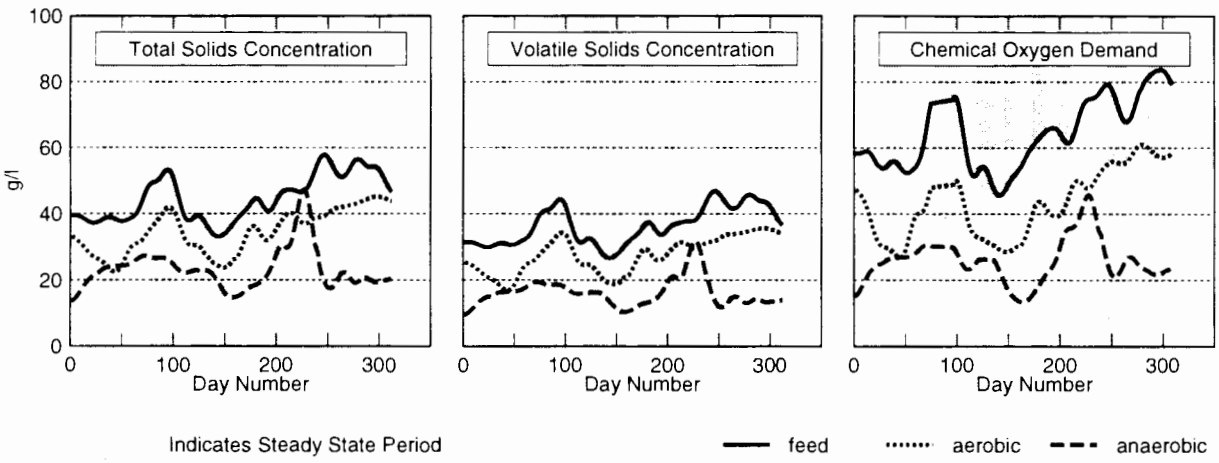

Figure 4.7 Graphical Representation of the Variation in Total Solids, Volatile Solids, and COD Concentrations for the Athlone Dual Digestion Process (Smoothed Data) During Phase I of the Investigation

The average rate of VS destruction in the aerobic reactor over the full 312 day evaluation period is given by the following equation:

$$M(VS)_{dest} = M(VS)_{in} - M(VS)_{out} = \sum_{j=1}^{312} Q(SL)_{in_j} \frac{(VS_{in} - VS_{out})_j}{24 \times 312} \quad \text{...kg(VS)/h} \quad (4.3)$$

where:

$M(VS)_{dest}$ = Rate of volatile solids destruction in the aerobic reactor (kg(VS)/d)

---

[2] At the start of the evaluation period, it was decided to operate the digester at the shortest possible retention time to test the conditioning effects of aerobic pre-treatment (see Section 3.1.10). It soon became apparent, however, that even at the minimum operating liquid level the capacity of the digester was too large to allow short retention times (10 to 15 days) to be tested. Because the gas mixing was poor and inefficient at the low liquid level (1100m^3 volume), from day 30 onwards, the digester was operated at the maximum operating liquid level (1800m^3 volume).

$M(VS)_{in,r}$ = Rate of volatile solids entering and leaving the reactor respectively (kg(VS)/d)
$Q(SL)_{in}$ = Influent sludge flow rate to the reactor (m^3/d)
$VS_{in,r}$ = VS concentration in the influent and effluent sludge stream (kg(VS)/m^3)
j = Denotes the day number of the evaluation period (from day j=1 to day j=312)

A similar equation can be written for the rate of COD removal. Employing Eq 4.3 the average rates of volatile solids destruction and COD removal over the full 312 day evaluation period are given in Table 4.13 below. On average 16.2 kg(VS) and 31.5 kg(COD) were destroyed per hour in the aerobic reactor, which is 25% and 33% respectively of the average VS and COD sludge flow rate through the system (64.8kg(VS)/h and 95.5kg(COD)/h).

Table 4.13 The Average Rate of Volatile Solids and COD Reduction in the Aerobic Reactor During Phase I: Oxygenation with Air Only

Parameter		Mean
Rate of Volatile Solids Destruction	kg(VS)/h	16.2
Rate of COD removal	kg(COD)/h	31.5

An indication of the variation in the percentage reduction in the VS and COD mass in the aerobic reactor through the evaluation period is graphically illustrated in Figure 4.8. As with the graphical presentation of the change in VS and COD concentrations, the data has been smoothed prior to plotting to eliminate any short term variations. The average percentage removal for each of the parameters is given in Table 4.14 below.

Table 4.14 The Percentage Reduction in TS, VS, and COD Across Each Stage in the Dual Digestion Process During Phase I: Oxygenation with Air

Treatment Stage	TS (%)	VS (%)	COD (%)
Aerobic	23	25	33
Anaerobic	32	40	40
Overall	48	56	59

From Table 4.14 it can be seen that an average 25% VS removal was achieved in the aerobic reactor. This is quite a high figure indicating that a considerable degree of stabilisation took place in the aerobic reactor. In the Milnerton pure oxygen reactor, the VS removal was very low (1.3%). The difference may be a consequence of the difference

in retention times which in the Milnerton reactor was 1.25 days compared with an average 4.4 days in this (Athlone) aerobic reactor. This statement could not be supported from the results of the phase II investigation during which the aerobic reactor was operated at 1 to 2 days retention time with pure oxygen supplementation (Chapter 7) - the %VS removal here was also higher than at Milnerton i.e. 20% cf. 1.2% (Section 7.4.1). The longer retention time and higher VS removal allows some of the VS destruction rate concepts developed for the ATAD system to be applied to the air oxygenated aerobic reactor in dual digestion (see Section 4.4.2 below, Section 3.10 and Chapter 5)

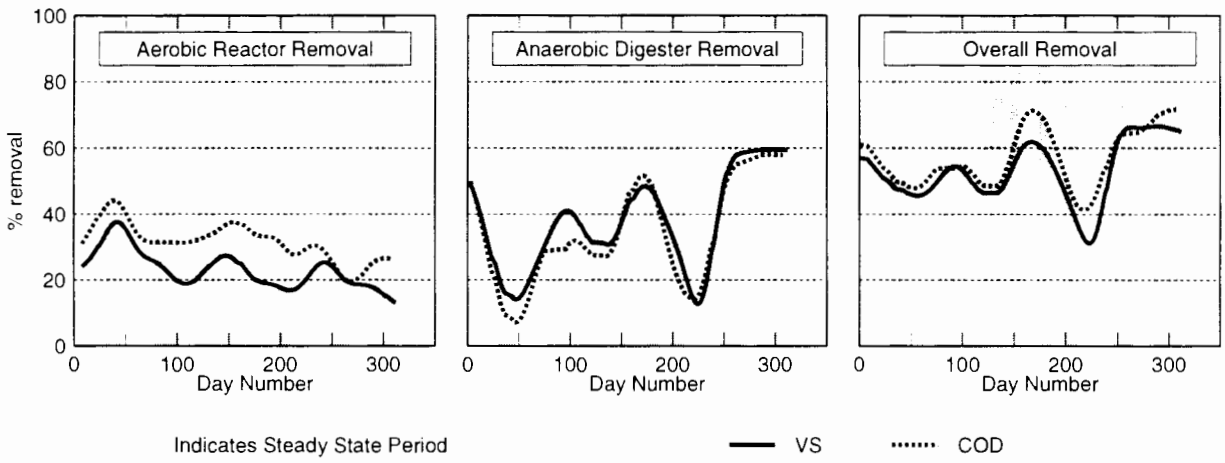

Figure 4.8 Graphical Representation of the Variation in Removal Efficiency for Volatile Solids and COD for each stage in the Athlone Duel Digestion Process (Smoothed Data). Phase I, Oxygenation is with Air Only.

4.4.2 Relationship between Biological Heat,Oxygen Demand and VS Destruction

The biological aerobic treatment of sewage sludges produces heat through the oxidative degradation of the organics in the sludge. It is reasonable therefore to expect that there will be a stoichiometric relationship between the rate of biological heat generation, the rate of volatile solids destruction, and the rate of oxygen utilisation. However, while this expectation may seem reasonable it was not met in the Milnerton pure oxygen aerobic reactor operating at 1.25 to 3 days. This expectation is met for ATAD's which operate at long (8-10d) retention times. Because the Athlone aerobic reactor was operated at around 4.5 days retention time, it is possible that the stoichiometric relationships for the ATAD systems may apply to it.

In Section 4.3.8 above, the biological heating rate H_h was quantified in terms of the oxygen utilisation rate ***OUR*** via the specific heat yield coefficient Y_h viz:

$$Y_h = 12.8 \text{ MJ/kg}(O_2)$$

The average oxygen utilisation rate $\overline{OUR}$ measured across the full period of phase I was 0.15 kg(O_2)/m^3.h. The $\overline{OUR}$ is volume specific, multiplication by the aerobic reactor process volume (V_p = 184m^3) gives the average process mass oxygen utilisation rate, viz:

$$\overline{M}(O_2)_{ut} = \mathbf{27.6\ kg(O_2)/h}$$

The average biological heating rate $\overline{H_b}$ during phase I is calculated (refer Eq 3.4) at:

$$\overline{H_b} = \mathbf{353\ MJ/h}$$

The average rate of volatile solids destruction $\overline{M}(VS)_{dest}$ during phase I (from Table 4.13) is as follows:

$$\overline{M}(VS)_{dest} = \mathbf{16.2\ kg(VS)/h}$$

The quantity of biological heat which is generated per unit mass of volatile solids destroyed $Y_{h(VS)}$ (which is the specific heat yield coefficient in terms of VS destroyed as apposed to the mass of oxygen utilised) is calculated as follows:

$$Y_{h(VS)} = \frac{H_b}{M(VS)_{dest}} = \frac{353}{16.2} \quad \text{...MJ/kg(VS)}$$

which yields:

$$Y_{h(VS)} = \mathbf{21.8\ MJ/kg(VS)}$$

The above value for the (volatile solid) heat yield coefficient $Y_{h(VS)}$ compares **very** favourably with the 21 MJ/kg(VS) obtained by Andrews and Kambhu (1971) for ATAD systems applications.

The final relationship between the three parameters is the rate of oxygen utilisation per mass of volatile solids destroyed f_{ovs}, viz:

$$f_{ovs} = \frac{Y_{h(VS)}}{Y_h} = \frac{21.8}{12.8} \quad \text{...kg(O}_2\text{)/kg(VS)} \quad (4.5)$$

which yields:

$$f_{ovs} = \mathbf{1.7\ kg(O_2)/kg(VS)}$$

The calculated figure for the oxygen utilised per VS destroyed (f_{ovs}) is higher than the theoretical COD/VS ratio (f_{cv}) for a typical wastewater sludge (usually accepted to be 1.42 kg(COD)/kg(VS)); however it is in agreement with the COD/VS ratio measured for the feed sludge to the aerobic reactor (1.7kg(COD)/kg(VS)). The ratio f_{ovs} of the mass of oxygen utilised to the mass of volatile solids destroyed is important in that it will be used later in Chapter 5, in the modelling of the aerobic reactor performance to predict the degree of VS removal. The fact that there is good agreement with both Messenger *et al* (1992) and Andrews and Kambhu (1971) regarding the stoichiometry of biological heat generation in terms of oxygen utilised and VS destroyed respectively, gives confidence that the measurements made on the reactor are correct. Furthermore, by comparing the input and output carbon mass flow rates via the influent and effluent sludge (accepted as $C_5H_7O_2H$) and gas streams, a carbon balance of >98% was obtained (see Appendix 6). Consequently, calculations and models based on these values are sufficiently accurate for application to design.

4.4.3 Aerobic Conditioning

Aerobic conditioning can be defined as the effect of aerobic pre-treatment on the sludge which subsequently reduces the minimum required retention time in the anaerobic digestion stage. This is brought about mainly as a result in:

- an increase in alkalinity and pH as a result of ammonification
- partial solublisation of particulate organic matter

An increase in pH and alkalinity as a result of ammonification after aerobic pre-treatment was manifest throughout the evaluation period and the average values of the pH, bicarbonate alkalinity and ammonia concentration of the feed and aerobic reactor sludges are given in Table 4.15; The smoothed variation in the pH, bicarbonate alkalinity, conductivity and ammonia concentrations of the feed, aerobic and anaerobic sludges throughout the evaluation period is shown graphically in Figure 4.9.

Table 4.15 The Average pH, Ammonia, and Alkalinity Concentrations in the Feed Sludge and Aerobic Sludge During Phase I

Sludge Type	pH		Ammonia Conc. mg(N)/ℓ		Bicarb Alk. $mg(CaCO_3)/\ell$	
	mean	range	mean	range	mean	range
raw feed	5.4	4.8-6.1	113	49-237	40	0-460
Aerobic	7.4	6.1-8.1	365	42-697	820	0-1460

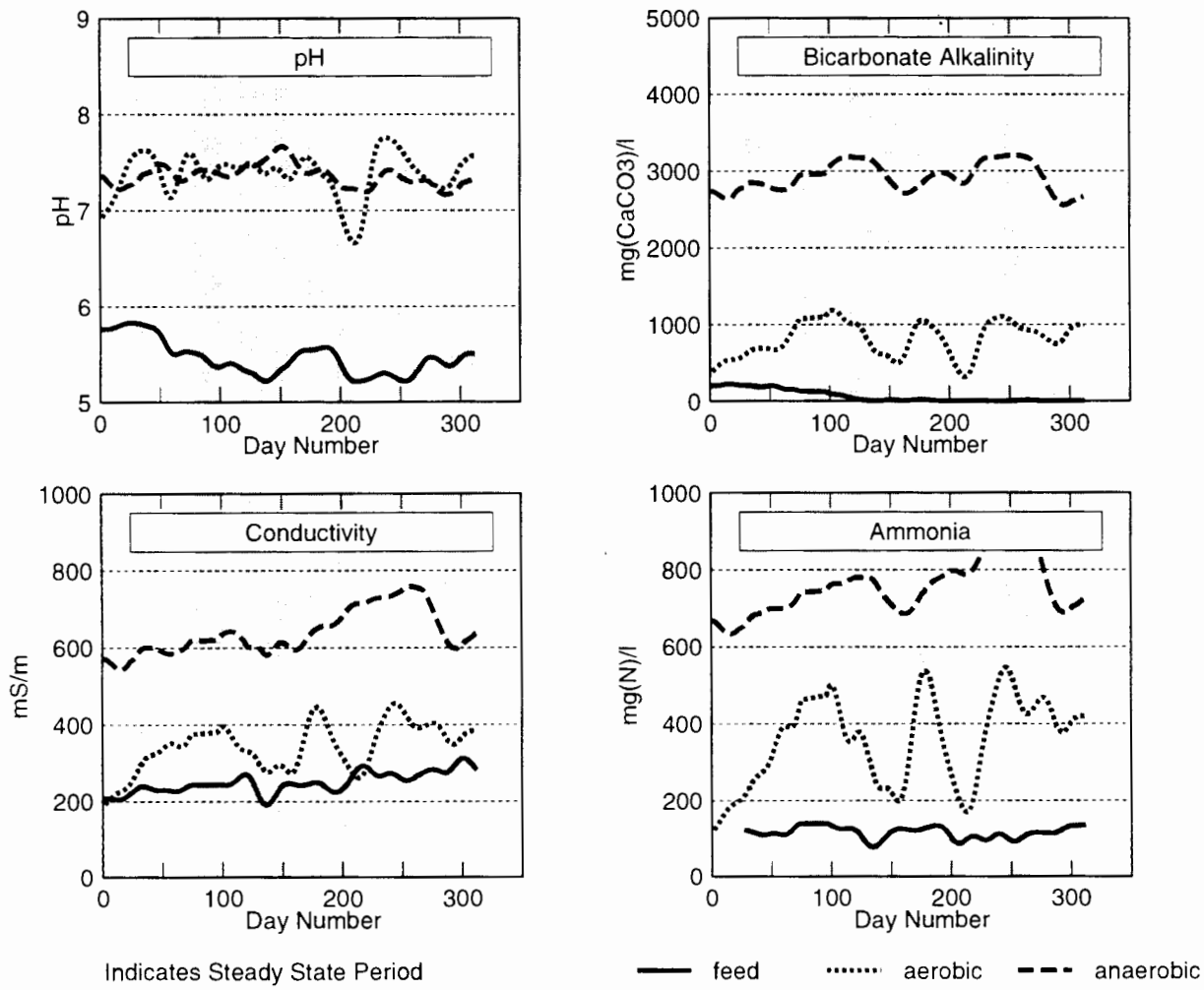

Figure 4.9 Graphical Representation of the Variation in pH, Alkalinity, Conductivity, and Ammonium Concentrations for the Athlone Duel Digestion Process (Smoothed Data): Results for Phase I

Conditioning of the feed sludge through the partial solublisation of particulate organic matter in the aerobic reactor was not evaluated. In order to carry out this evaluation, the soluble COD would have had to be determined on both the feed and aerobic sludges; an increase in the soluble COD across the aerobic reactor would have been indicative of the solublisation of particulate organic matter taking place within the reactor. However, because it was not practical to separate the liquid and solid phases in the samples from the aerobic reactor to measure the soluble COD (as was found at Milnerton also, Messenger *et al*, 1992), the solublisation of particulate organic matter could not be monitored. Conditioning, therefore was assessed in terms of pH, conductivity, ammonium and alkalinity increases. In this regard, Table 4.15 and Fig 4.9 show that the aerobic reactor sludge is a much more suitable feed for anaerobic digestion because it has a neutral pH and a high bicarbonate alkalinity.

As the anaerobic digester operating volume was fixed at 1800m^3 (excluding the short period of time immediately after start-up (± day 1 to 50) when the digester was operated at the minimum operating level of 1100m^3), the shortest sludge retention time achieved during the study was 30 days (period 6). Consequently, it was not possible to verify, at full scale, the claim made for the dual digestion process that aerobic pre-treatment reduces the minimum required anaerobic retention time. This claim was tested at laboratory scale (see Section 4.5 below for results).

4.4.4 Bacteriological Quality

The degree of disinfection provided by the dual digestion system was monitored by determining faecal coliform counts and *Ascaris* ova viability across at each stage of the system. The results observed over the 312 day evaluation period are listed in Appendix 2. The median faecal coliform values were abstracted from this Appendix and are given in Table 4.16 below. The distribution in values for the feed, aerobic and anaerobic sludges is shown in Figure 4.10 below.

Table 4.16 Summary of Faecal Coliform Analysis for Phase I

Faecal Coliforms/100mℓ - Medians		
Raw	Aerobic	Anaerobic
9.3×10^8	7.5×10^5	4.3×10^4

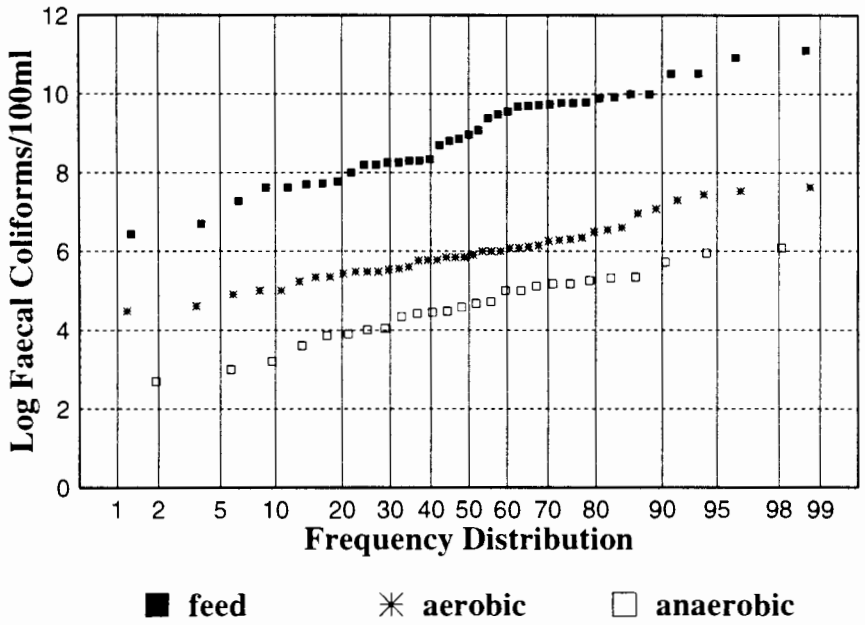

Figure 4.10 Frequency Distribution of Measured Faecal Coliform Counts: Phase I

The parameter used in the evaluation to assess the degree of disinfection achieved by the system was measurement of the inactivation of *Ascaris* ova. The percentage viable *Ascaris* ova recorded in the sludge after aerobic treatment as a function of reactor temperature is shown in Figure 4.11 below:

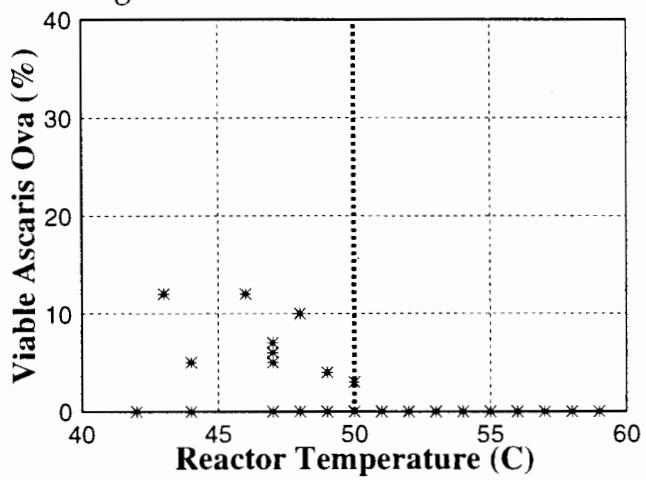

Figure 4.11 Viable Ascaris Ova in Aerobic Reactor Sludge versus Reactor Temperature: Phase I

From Figure 4.11 it can be seen that disinfection in terms of inactivation (zero viable) of Ascaris ova occurred only when the aerobic reactor temperature was in excess of 50°C. This fact is further highlighted in Table 4.17 below. Mean values for the raw feed sludge and anaerobic digester sludge are also quoted.

Table 4.17 Summary of *Ascaris* Ova Analysis for Phase I

Ascaris Ova/g dry sludge (% Viable) - Mean Values			
Raw	Aerobic		Anaerobic
	Temp <50°C	Temp >50°C	
429 (45%)	415 (3%)	545 (0%)	758 (12%)

At no point during the evaluation period did the sludge from the anaerobic digester become free from viable *Ascaris* ova. This could be attributed to the fact that at the start of the evaluation period all the sludge in the anaerobic digester (1100m^3) was obtained by seeding with sludge from the existing mesophilic digestion plant, and as such was contaminated. Also when the digester volume was increased to 1800m^3, a further 700m^3 of "contaminated" conventional digested sludge was added into the digester. Due to the long retention time in the digester it was not possible for the viable ova to be washed out from the digester.

The significance of 50°C in terms of inactivating the *Ascaris* ova is used in the modelling described in Chapter 5. However it is recognised that this temperature is below normal sludge pasteurisation specifications (DNH&PD, 1991).[3] The temperature of 50°C is accepted to be the minimum desired temperature for operation of the aerobic reactor and as such theoretical feed volumes are calculated on achieving this temperature..

4.4.5 The Dewaterability of the Final Sludge

The dewaterability of the final treated sludge was measured by means of the specific resistance to filtration (SRF) test (Smollen, 1986 and Swanwick *et al*, 1962). A total of ten tests were performed during the 312 day evaluation period, the results of which are listed in Appendix 2. These results are summarised in Table 4.18 and comparison is made with values obtained after; (1) conventional mesophilic anaerobic digestion at the the Athlone Wastewater Treatment Plant and (2) Dual Digestion with pure oxygen at Milnerton (Messenger *et al*, 1992).

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[3] i.e. 70°C for >30 minutes. It is interesting to note that the UK Dept. of Environment (1990) accept also >55°C for >4h or appropriate intermediate conditions between this and >70°C for >30 minutes, subject to the condition of subsequent primary mesophilic anaerobic digestion.

Table 4.18 Comparison of the Dewaterability Characteristics of Sludges from the Dual Digestion Process, both with Air and with Oxygen, and the Anaerobic Digestion Process

Sludge Treatment	SRF (median) m/kg x 10^{12}	No of Tests	Sludge Temp Achieved
Dual Digestion using Air	368	10	50°C
Dual Digestion using Oxygen	507	6	65°C
Mesophilic Anaerobic Digestion	218	10	37°C

By comparing the SRF data in Table 4.18 for each of the different processes, it can be concluded that the dewaterability of the final sludge from the dual digestion plant at Athlone (where the aerobic reactor is oxygenated with air) is not significantly different to the dewaterability of the final sludge from either the dual digestion plant at Milnerton (where the aerobic reactor is oxygenated with pure oxygen) or with that from the convention mesophilic anaerobic digestion plant at Athlone. It is possible though, considering, the maximum temperature reached in the three digestion systems, that the SRF may be dependent on this, increasing as the temperature increases. While this link with temperature may exist, it is nevertheless clear that all three sludges dewater rather poorly (c.f. Smollen, 1986).

4.5 LABORATORY SCALE STUDY OF THE EFFECTS OF AEROBIC CONDITIONING

To determine whether or not sludge conditioning in the autothermal thermophilic aerobic reactor is able to reduce the retention time in the anaerobic digester below that normally associated with conventional anaerobic digestion (15days), two laboratory scale digesters (Digester 1 and Digester 2), each with an operating capacity of 6l, were operated for a period of two months at a sludge retention time of 8 days. Both were controlled at 37°C by standing them in a hot water temperature bath.

Digester 1 was fed with raw primary sludge from the gravity thickener, whilst Digester 2 was fed with sludge taken from the aerobic reactor. These feed sludges were obtained from the raw primary and aerobic reactor composited sludge samples collected for analysis over a 24 hour period from the full scale dual digestion system. At the start of the evaluation period, both digesters were filled with active sludge from the anaerobic

digester. A full description of the units and the mode of operation is given in Section 2.5 above. All the results from the study are listed in Appendix 3.

In Digester 1 (fed with raw primary sludge) the pH level dropped soon after start-up and a gradual build up of solids caused pumping problems with the desludging pump. By day 27 the pH had dropped below 6.0, indicating that digester 1 had failed and the study on it was terminated.

Digester 2 (fed with aerobic reactor sludge) remained in a stable condition throughout the two month study period. Average data for the final three weeks of operation of Digester 2 are presented in Table 4.19 below. The change in the main chemical parameters in Digester 2 over the 50 day study period are given in Figure 4.12.

Table 4.19 Average Chemical Data from Laboratory Scale Digester No 2

Sludge Type	TS	VS	Ammonia	pH	VAA	BA
	g/ℓ	g/ℓ	mg(N)/ℓ	-	mg($CaCO_3$)/ℓ	mg($CaCO_3$)/ℓ
Feed (ex Aerobic)	43.3	34.7	434	7.4	670	870
Final Anaerobic	25.0	18.7	860	7.1	970	2100
Increase/Decrease	-18.3	-16.0	+426	-0.3	+300	+1230
Percentage Change	-42	-46	+97	-	+45	+141

The results of the laboratory scale studies indicate that, by undergoing pre-treatment in the aerobic reactor, the primary sludge was able to be anaerobically digested at a shorter retention time (in this case 8 days) than would otherwise have been possible, while maintaining a high degree of VS removal (46%). The alkalinity and the pH of the feed sludge to Digester 2 were significantly higher (as a result of aerobic pre-treatment) in comparison to the alkalinity and pH of the feed sludge to Digester 1 (which failed soon after start-up). Consequently the impact on the buffer capacity of the sludge in Digester 2 would have been a lot lower, and this fact would account for the stability of this particular digester. However because of the length of the retention time in the aerobic reactor (4d), a relatively high degree of volatile solids destruction takes place (about 25%). Therefore, the stability and successful operation of Digester 2 may well be a consequence of the fact that the sludge from the aerobic reactor was already partially stabilised, and not so much that it has become conditioned i.e. that the alkalinity has increased and that partial solublisation of particulate organics has taken place.

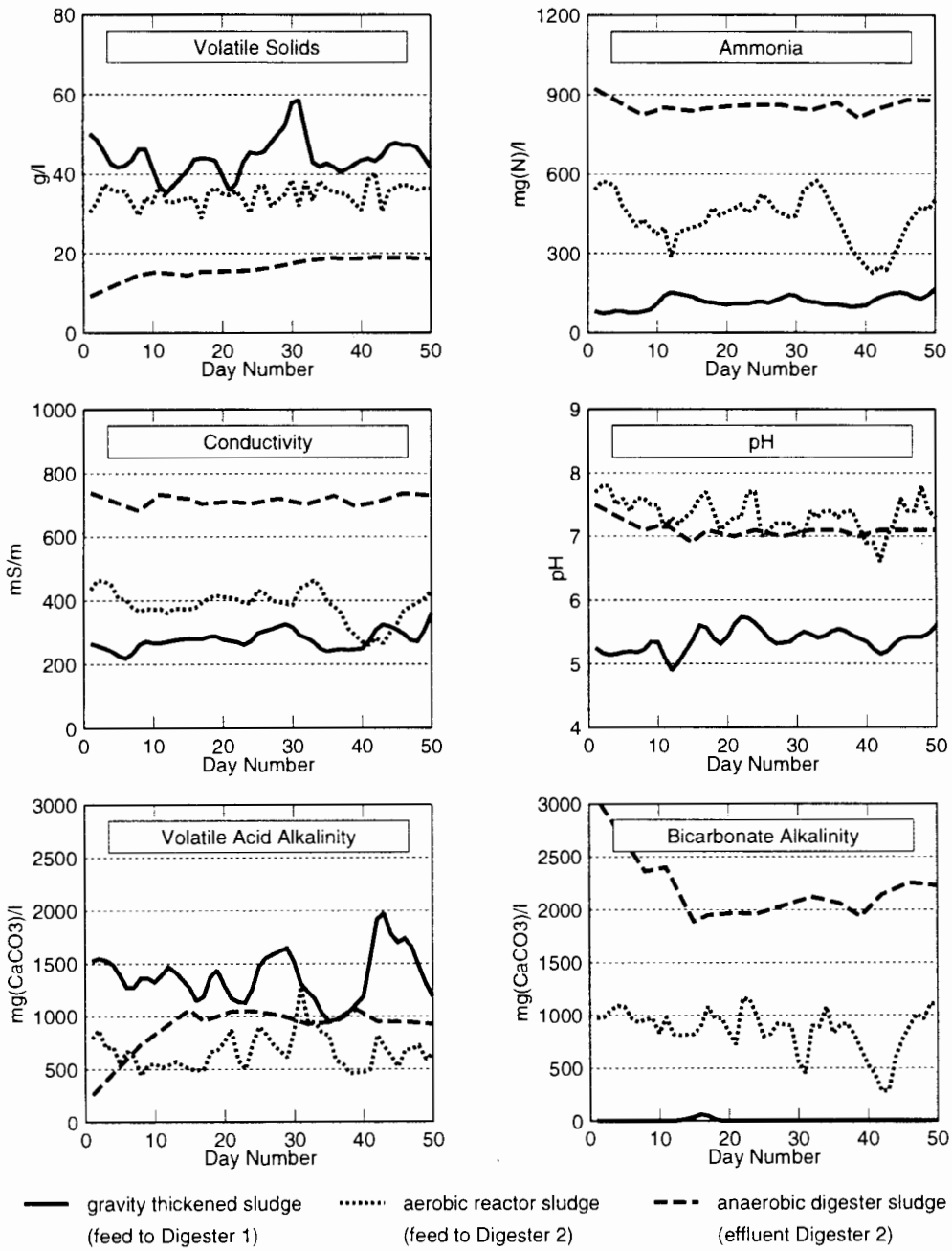

Figure 4.12 Graphical Representation of the Variation in the Main Sludge Chemical Characteristics for the Laboratory Scale Anaerobic Digester 2 (Smoothed Data): Conducted during Phase I

The VS removal in the laboratory scale digester at an 8 day retention time was 46% (Table 4.19). This is considered to be a very good VS removal and well above that considered necessary for sludge stabilisation (>38%, Heidman, 1989).[4] It is interesting to note that the VS removal in the laboratory digester was greater than in the full scale

[4] This is not the only criterion for assessing sludge stability; a specific *OUR* < 1.0mg(O)/g(TS).h is usually applied

digester (40%), fed with the same aerobic reactor sludge, which operated at an average retention time of 42 days. The greater degree of VS destruction in the laboratory scale digester may be attributed to the higher operating temperature (37°C compared to 31°C) and more efficient mixing in the laboratory scale digester. The results from the laboratory scale study supports the claim made by proponents of the dual digestion system that the anaerobic retention time can be reduced as a consequence of the sludge heat pre-treatment in the aerobic reactor. In a similar laboratory study, Izzett *et al* (1992) demonstrated that heat treated (65°C for 24h) and raw primary sludge could both be successfully anaerobically digested at a retention time of 7 days; at shorter retention times both systems failed (turned sour). The %VS removals achieved were virtually identical and decreased from 53% at a 20 day retention time to 42% at 7 days. Although in this study the sludge was not heated autothermally (i.e. not biologically but electrically), it also indicated that anaerobic digesters could be operated at significantly shorter retention times than is conventionally practical (20 to 30 days). This fact is further evidenced at the Cape Flats (150Ml/d) wastewater treatment plant, where the raw primary sludge is steam pasteurised at 70°C for 30 minutes prior to stabilisation by conventional mesophilic anaerobic digestion (Morrison, 1986). In the period 1991-94, the anaerobic digesters have successfully operated at an average retention time of 10 days achieving approximately 48% VS removal.

In evaluating the merits of a particular process, it is important to distinguish between process stability (i.e. where the main process parameters remain relatively constant under steady state conditions) and product stability (i.e. where specific process parameters meet with certain criteria, viz %VS reduction >38%, specific ***OUR*** <1.0 $mg(O_2)/g(TSS).h$). For example, Drnevich and Matsch (1978) found that, in a bench scale dual digestion system, the anaerobic digesters were able to be operated at a retention time of 3 days (pH = 7.0; Temp = 35°C). The %VS removal in the digester at this retention time was 16% with an overall system VS removal of 28%. Whilst the anaerobic process could be defined as stable at this short retention time (3 days), the product (in terms of >38%VS reduction criterion) is considered unstable.

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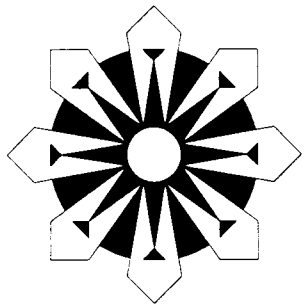

CHAPTER 5

MODELLING AEROBIC REACTOR PERFORMANCE: PHASE I OXYGENATION USING AIR

5.1 INTRODUCTION

5.1.1 Application of the Steady State Heat Balance

One of the major benefits derived from performing a steady state heat balance on the aerobic reactor is that, once calibrated with experimental data (in this instance obtained from the 312 day evaluation period of phase I), it can be used to compile design charts which enable the performance of the aerobic reactor to be predicted under a variety of different operating conditions. This subsequently allows for:

- Optimisation of system operation
- Evaluation of the viability of the system
- Assistance in future design

The initial objective of performing the steady state heat balance for each period of stable operation during phase I (see Section 4.3.7) was to determine the specific heat yield coefficient Y_h (see Section 4.3.8)[1]. Inspite of the different steady state operating conditions established during phase I, the specific heat yield coefficient Y_h (calculated from the heat and oxygen mass balances) remained relatively constant, ranging between 12.3 and 13.2 MJ/kg(O_2) with an average of 12.8 MJ/kg(O_2) (see Table 4.10). This indicates that for the range of operating conditions during phase I, the heat and oxygen mass balance equations are sufficiently general to allow the system to be accurately modelled. The ranges of each of the main operating parameters over the 8 different steady state periods of phase I are given in Table 5.1 below:

[1] The specific heat yield coefficient Y_h is defined as the quantity of biological heat generated per unit mass of oxygen utilised by the aerobic organisms (MJ/kg(O_2)).

Table 5.1 The Ranges of Each of the Main Operating Parameters over the Eight Different Steady State Periods During Phase I: Oxygenation with Air

Symbol	Parameter	Units	Min	Max
$T(SL)_{in}$	Influent Sludge Temperature	°C	16	23
$T(AIR)_{in}$	Influent Air Temperature	°C	15	25
$Q(SL)_{in}$	Influent Sludge Flow Rate	m^3/d	28	60
$Q(AIR)_{in}$	Influent Air Flow Rate	m^3(STP)/h	230	1200
$T(SL)_r$	Aerobic Reactor Temperature	°C	44	56

Having (1) calibrated the heat and oxygen mass balance equations by determining such constants as the specific heat yield Y_h and respiration quotient Y_{CO2} (see Section 4.3.6), and (2) verified the generality of these equations by demonstrating the reproducibility of the values of these constants under different operating conditions, it is permissible to employ the same heat and oxygen mass balance equations for design. This is done by accepting the values of the measured constants and calculating process parameters from specified operating conditions. By plotting in a design chart a particular process design parameter against a selected varying operating condition while the other parameters and conditions remain constant, allows the effect of the operating condition on the design parameter to be evaluated. The development of such design charts, where oxygenation of the aerobic reactor is with air alone, is the principal objective of this Chapter[2].

5.1.2 Summary of the Design Models: Oxygenation with Air

Design charts and equations are derived in this Chapter for predicting operating conditions in the aerobic reactor where oxygenation is with air alone, concluding with a prediction of oxygen supplementation rates required for phase II of the investigation (where oxygenation is with air + pure oxygen). The models are as follows:

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[2] Modelling of the aerobic reactor, where oxygenation is with both air + pure oxygen, is dealt with in Chapter 8 below.

Model (1) Oxygenation with Air: Non-Foaming Conditions (Section 5.2)

- Part one of the first model predicts the maximum allowable feed sludge loading rate and retention time at different ambient temperatures to sustain a reactor temperature of 50°C under non-foaming conditions..

- Part two of the first model predicts the minimum required feed solids concentration at different retention times to ensure that the aerobic reactor does not become substrate limited. It also predicts the percentage volatile solids destruction and the increase in the ammonium ion concentration as a result of the breakdown of volatile solids in the aerobic reactor versus retention time.

Model (2) Oygenation with Air: Foaming Conditions (Section 5.3)

- At the onset of foaming, the air flow rate is reduced to prevent foam spillage (refer Section 4.3.5). The second model predicts the change in reactor temperature with airflow rate at constant ambient temperature under foaming conditions.

- Due to the improvement in oxygenation characteristics with foaming (refer Section 4.3.5), an increase in the maximum feed sludge loading rate is possible whilst maintaining a reactor temperature of 50°C. This second model predicts the change in the maximum feed sludge loading rate with airflow rate at constant ambient temperature under foaming conditions..

Model (3) Prediction of Pure Oxygen Supplementation Rates (Section 5.4)

- The third model predicts the quantity of pure oxygen required to maintain operating temperatures above 50°C at differing feed loading rates and differing ambient temperatures under both foaming and non-foaming conditions. This model was used to correctly size the pure oxygen injection equipment for phase II.

The derivation for each of the above models is described below. The formulae for calculating each parameter are given and design charts are presented for ease of use which indicate conditions for summer and winter operation with air oxygenation only.

5.2 MODELLING NON-FOAMING CONDITIONS

5.2.1 Observations Made During the Evaluation Period

Throughout phase I (when oxygenation was with air alone) no significant foam[3] layer was observed either during the period between start-up at ambient temperature to the point where the reactor temperature reached 50°C, or on subsequent occasions when the reactor temperature dropped below 50°C i.e.

> **For oxygenation with air alone, significant foaming only occurred in the aerobic reactor when the reactor temperature was in excess of 50°C**
>
> ...(see Section 4.3.10)

Consequently, in order predict the operating conditions required to sustain a reactor temperature of 50°C the process must be modelled under non-foaming conditions i.e. the objective is first to reach thermophilic temperatures under non-foaming conditions, and then if foaming occurs accept this as a bonus i.e. higher temperatures will be achieved giving a higher degree of disinfection and providing greater sensible heat to the anaerobic digester.

During phase I it was found that with the dry influent air flow rate set at **760 m^3(STP)/h** (The maximum delivery rate from one compressor), and **no foam** present on the surface of the sludge, the oxygen utilisation rate remained relatively constant at **0.15kg(O_2)/m^3.h** (see Section 4.3.3). It is accepted that when the air stream exits from the surface of the sludge it is at the same temperature as that of the sludge (which for modelling is taken to be **50°C**) and is fully saturated with water vapour (i.e equilibrium has been reached between the liquid and gas phases in terms of mass and heat transfer). The effluent air stream then undergoes some degree of cooling (estimated at **3°C**) in the vacant reactor headspace and exits the reactor at **47°C**. At 47°C, the saturation humidity U_{out} of the effluent air stream is calculated to be **0.074kg(H_2O)/kg(AIR)** (Eq 3.124).

For simplicity, both the influent air stream temperature and the influent feed sludge temperature, are accepted as equal and denoted by the term T_{amb} (ambient temperature). For ease of calculation, the influent air stream humidity is accepted to vary linearly with the ambient temperature (see Figure 5.1 below).

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[3] The foam layer was considered to be significant when the depth exceeded 1m for more than 10 days (see Figure 4.6).

$$U_{in} = 0.00058\,T_{amb} - 0.0020 \quad \text{...kg}(H_2O)/\text{kg(AIR)} \quad (5.1)$$

Figure 5.1 below shows the variation of the influent air stream humidity with air stream temperature (the air stream is presumed fully saturated with water vapour) and compares the linear approximation Eq 5.1 (dotted line) with the theoretically calculated humidity Eq 3.124 (solid line).

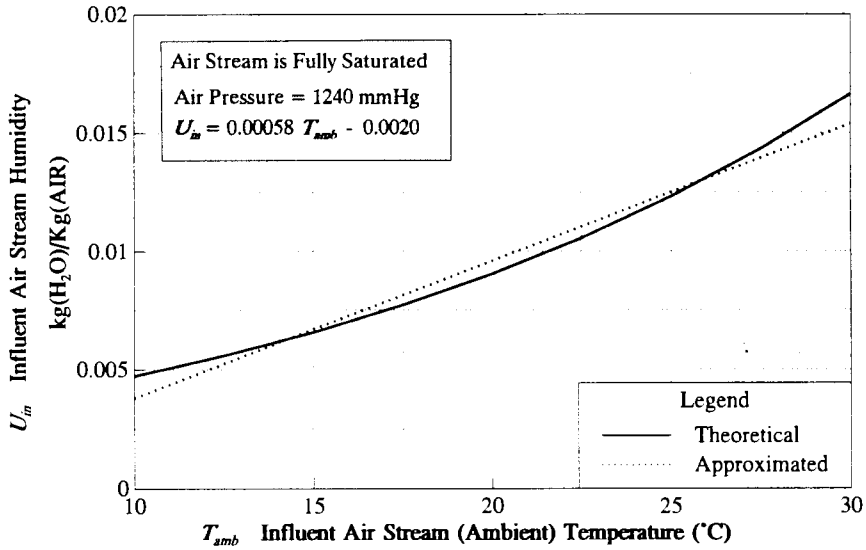

Figure 5.1 The Variation of Theoretical and Approximated Influent Air Stream Humidities (kg(H_2O)/kg(AIR)) with Air Stream Temperature.

5.2.2 Maximum Sludge Loading Rate

The first model for non-foaming conditions predicts the maximum sludge loading rate (minimum sludge retention time) which will enable the Athlone aerobic reactor to operate at a temperature of 50°C whilst oxygenation is with air alone. The sludge loading rate (aerobic reactor retention time) is expressed as as a function of the ambient temperature, thereby enabling maximum loading rates to be predicted for both the winter and summer months. The prediction is based on the solution of the Steady State Heat Balance arranged in terms of the retention time (R_h). It is assumed that there is no limitation on available substrate i.e that the reactor operates under oxygen limiting conditions. In the derivation each heat term in the balance is described separately.

The Biological Heating Rate

$$H_b = Y_h . V_p . OUR \quad \text{...MJ/h} \quad (5.2)$$

where:

Y_h	= Specific Heat Yield	12.8	$MJ/kg(O_2)$
V_p	= Aerobic reactor process volume	184	m^3
OUR	= Oxygen utilisation rate	0.15	$kg(O_2)/m^3.h$

$$H_b = 12.8 \times 184 \times 0.15 = 353 \quad \text{...MJ/h} \quad (5.3)$$

The Mechanical Heat Input Rate

$$H_m = 0.0036 \times f_{mech} . \cos\phi . \sqrt{3} . V . I \quad \text{...MJ/h} \quad (5.4)$$

where:

f_{mech}	= Fraction of electrical energy converted to heat	0.85	
V	= Phase voltage of the pump motor	380	volts
I	= Phase current drawn by the pump motor	19	amps
$cos\Phi$	= Power factor for the pump motor	0.82	
0.0036	= Conversion factor from W to MJ/h	0.0036	MJ/W.h

$$H_m = 0.0036 \times 0.85 \times 0.82 \times 1.732 \times 380 \times 19 = 31 \quad \text{...MJ/h} \quad (5.5)$$

The Effluent Gas Vapour Heat Loss Rate

$$H_v = L_v^{50^\circ C} . \rho(AIR) . Q(AIR)_{in} . (U_{out} - U_{in}) \quad \text{...MJ/h} \quad (5.6)$$

where:

$L_v^{50^\circ C}$	= Latent heat of vaporisation of water at 50°C	2.38	$MJ/kg(H_2O)$
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	$kg(AIR)/m^3$
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	760	$m^3(STP)/h$
U_{out}	= Effluent air stream humidity (at 47°C)	0.074	$kg(H_2O)/kg(AIR)$
U_{in}	= Influent air stream humidity (function of T_{amb})	$f(T_{amb})$	$kg(H_2O)/kg(AIR)$

$$H_v = 2.38 \times 1.21 \times 760 \left(0.074 - (0.00058 T_{amb} - 0.0020)\right) = 166 - 1.26 T_{amb} \quad \text{...MJ/h} \quad (5.7)$$

The Effluent Gas Sensible Heat Loss Rate

$$H_g = \rho(AIR) . Q(AIR)_{in} . \left(C_p^{20^\circ C}(AIR) . (T(AIR)_r - T(AIR)_{in}) + U_{out} . C_p^{50^\circ C}(H_2O)_{vap} . T(AIR)_r\right) \quad \text{...MJ/h} \quad (5.8)$$

where:

$\rho\ (AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	$kg(AIR)/m^3$
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	760	$m^3(STP)/h$
$C_p^{20^\circ C}$ (AIR)	= Heat capacity of dry air at STP	0.0010	MJ/kg(AIR)
$C_p^{50^\circ C}$ $(H_2O)_{vap}$	=Heat capacity of water vapour at 50°C	0.00187	$MJ/kg(H_2O)$
U_{out}	= Effluent air stream humidity (at 47°C)	0.074	$kg(H_2O)/kg(AIR)$
$T(AIR)_{in}$	= Influent air stream temperature	T_{amb}	°C
$T(AIR)_r$	= Effluent air stream temperature	47	°C

$$H_g = 1.21 \times 760 (0.0010 (47 - T_{amb}) + 0.074 \times 0.00187 \times 47) = 49 - 0.92 T_{amb} \quad \text{...MJ/h} \quad (5.9)$$

The Effluent Sludge Sensible Heat Loss Rate

$$H_s = C_p^{20^\circ C}(SL)\left(\left(\frac{Q(SL)_{in}}{24} - \frac{(\rho(AIR).Q(AIR)_{in}.U_{out}}{1000}\right).T(SL)_r - \frac{Q(SL)_{in}.T(SL)_{in}}{24}\right) \quad \text{...MJ/h} \quad (5.10)$$

where:

$C_p^{20^\circ C}(SL)$	= Heat capacity of sludge at 20°C	4.0	MJ/m^3
$Q(SL)_{in}$	= Influent sludge stream flow rate	Q_i^{max}	m^3/d
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.205	$kg(AIR)/m^3$
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	760	$m^3(STP)/h$
U_{out}	= Effluent air stream humidity (at 47°C)	0.074	$kg(H_2O)/kg(AIR)$
$T(SL)_{in}$	= Influent sludge stream temperature	T_{amb}	°C
$T(SL)_r$	= Effluent sludge stream temperature	50	°C
$T(AIR)_{in}$	= Effluent air stream temperature	47	°C

$$H_s = 4.0\left(\left(\frac{Q_i^{max}}{24} - \frac{1.205 \times 760 \times 0.074}{1000}\right)50 - \frac{Q_i^{max} T_{amb}}{24}\right) = 8.33Q_i^{max} - 14 - 0.167Q_i^{max}T_{amb} \quad \text{...MJ/h} \quad (5.11)$$

The Wall Heat Loss Rate

$$H_w = 0.82(T(SL)_r - T(SL)_d) + 0.49(T(SL)_r - T(AIR)_{in}) \quad \text{...MJ/h} \quad (5.12)$$

where:

$T(SL)_r$	= Aerobic reactor sludge temperature	50	°C
$T(AIR)_{in}$	= Ambient air temperature	T_{amb}	°C
$T(SL)_d$	= Anaerobic digester sludge temperature	$(50+T_{amb})/2$	°C

$$H_w = 0.82(50 - (50 + T_{amb})/2) + 0.49(50 - T_{amb}) = 45 - 0.9T_{amb} \quad \text{...MJ/h} \quad (5.13)$$

Combining each of the above heat terms in the steady state heat balance:

$$H_b + H_m = H_v + H_g + H_s + H_w \quad \text{...MJ/h} \quad (3.8)$$

yields:

$$353+31 = 166-1.26T_{amb}+49-0.92T_{amb}+8.33Q_i^{max}-14-0.167Q_i^{max}T_{amb}+45-0.9T_{amb} \quad \text{...MJ/h} \quad (5.14)$$

Rearranging in terms of Q_i^{max} yields:

$$Q_i^{max} = \frac{138 + 3.1T_{amb}}{8.33 - 0.167T_{amb}}$$

Equation 5.15 **The Maximum Sludge Flow Rate (m^3/d) which will allow a Reactor Temperature of 50°C at Different Ambient Temperatures: Oxygenation is with Air (Flow Rate 760 m^3 (STP)/h and No Foam is Present.**

The variation in the maximum sludge flow rate which will allow the aerobic reactor to reach an operating temperature of 50°C is illustrated in Figure 5.2. As the sludge flow

The variation in the maximum sludge flow rate which will allow the aerobic reactor to reach an operating temperature of 50°C is illustrated in Figure 5.2. As the sludge flow rate is linked to the sludge retention time ($R_h = V_r/Q_i$), the minimum retention time is also indicated. Note that this graph is valid for conditions when foam is not present, and the air compressor is operating at its maximum flow rate of 760 m^3 (STP)/h.

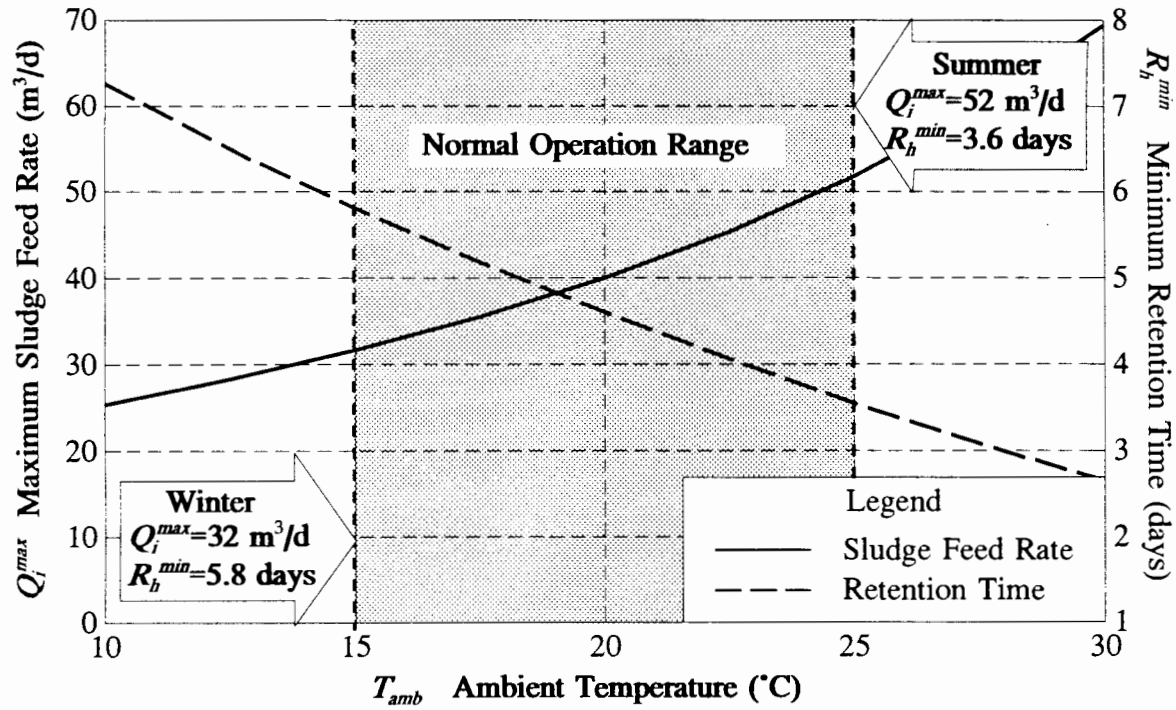

Figure 5.2 The Variation in the Maximum Sludge Feed Rate (Minimum Retention Time), which will allow a Reactor Temperature of 50°C, with Ambient Temperature: Oxygenation is with Air (Flow Rate 760 m^3 (STP)/h) and there is No Foam Present.

For design purposes, accepting an average ambient temperature of 25°C in summer and 15°C in winter, Table 5.2 below gives the maximum sludge flow rate (minimum retention time) for operation during these periods:

Table 5.2 Maximum Sludge Flow Rate (Minimum Sludge Retention Time) to Achieve a Reactor Temperature of 50°C during Summer and Winter: Oxygenation is with Air (Flow Rate 760 m^3(STP)/h) and No Foam is Present.

SEASON	T_{amb} °C	Q_i^{max} m^3/d	R_h d
Winter	15°	32	5.8
Average	20°	40	4.6
Summer	25°	52	3.6

5.2.3 Minimum Required Feed Sludge Concentration

The second part of model (1) for non-foaming conditions predicts the minimum feed sludge concentration required to ensure that substrate limitation does not take place in the aerobic reactor at different sludge retention times.

The first part of model (1) which predicted maximum sludge loading rates (Section 5.2.2), was based on the assumption that the reactor is operating under oxygen limiting conditions (which was the case during each of the steady state periods in the evaluation). However if there is either;

- an increase in the aerobic reactor retention time, or
- a decrease in the influent sludge solids concentration

then the reactor may become **substrate limited**. Under substrate limiting conditions the oxygen utilisation rate will fall below the oxygen transfer rate of the aeration device and high dissolved oxygen levels will be observed in the sludge. Substrate limitation in the reactor should be avoided because it reduces the oxygen utilisation rate *(OUR)* of the sludge which in turn reduces the biological heating rate (H_b). Any reduction in H_b will then cause a reduction in the aerobic reactor temperature.

In Chapter 3 kinetic equations were derived which describe the *OUR* (and volatile solids destruction) under both oxygen and substrate limiting conditions. At the point at which the reactor is at the transition between oxygen and substrate limiting conditions both sets of kinetics will be valid. The point of transition can therefore be determined by equating both sets of kinetics (see Section 3.11) viz:

Oxygen Limiting = Substrate Limiting

$$M(VS)_{dest} = \frac{V_p . OUR}{f_{ovs}} = \frac{k_d . f_{bio} . f_{vs} . TS_{in}^{min} . V_p}{24(1 + R_h . k_d)} \quad \text{...kg(VS)/h} \quad (3.221)$$

Which rearranged in terms of influent feed sludge total solids concentration yields:

$$TS_{in} = \frac{24 . OUR(1 + R_h . k_d)}{k_d . f_{ovs} . f_{bio} . f_{vs}} \quad \text{...kg(TS)/m}^3 \quad (3.223)$$

where:

TS_{in}	= Total solids concentration in the influent sludge stream	TS_{in}^{min}	kg(TS)/m^3
f_{ovs}	= Mass of VS destroyed per mass of oxygen utilised in the reactor	1.70	kg(VS)/kg(O_2)
f_{bio}	= Biodegradable fraction of the volatile solids in the influent sludge	0.60	kg/(BVS)/kg(VS)
f_{vs}	= Volatile solids fraction of the total solids in the influent sludge	0.81	kg(VS)/kg(TS)
R_h	= Retention time at which the reactor becomes substrate limited.	Rh	d
OUR	= Oxygen utilisation rate	0.15	kg(O_2)/m^3.h
k_d	= Volatile solid destruction rate coefficient at 50°C	0.25	d^{-1}

By equating the oxygen and substrate limiting condition equation as in Eq 3.221, a link between TS_{in} and R_h is established where TS_{in} is the minimum influent TS concentration (TS_{in}^{min}) to avoid substrate limitation at retention time R_h . As neither the biodegradable fraction of the volatile solids f_{bio} or the volatile solids destruction rate coefficient k_d were determined during the evaluation, the use of the above expression (Eq 3.223) is rather speculative. It is nevertheless still instructive to use it in order to give an indication of the influent sludge solids concentration required to prevent substrate limitation from taking place. A value of 0.60kg(BVS)/kg(VS) for f_{bio} is considered appropriate for the type of primary sludge produced at Athlone. A value of 0.25 d^{-1} is accepted for k_d at 50°C (taken from Fuggle and Spensley, 1985). Substitution of the appropriate values for the parameters in Eq 3.223 yields:

$$TS_{in}^{min} = \frac{24 \times 0.15 \times (1 + 0.25 \times R_h)}{0.25 \times 1.70 \times 0.60 \times 0.81} \quad \text{...kg(TS)/m}^3 \quad (5.16)$$

Which simplifies to:

$$TS_{in}^{min} = 4.4R_h + 17.4$$

Equation 5.17 **The Minimum Influent Sludge Solids Concentration Required (kg(TS)/m^3) to Prevent the Aerobic Reactor from becoming Substrate Limited as a function of the Retention Time (d): Valid for Non-Foaming conditions with the Air Flow Rate at Maximum and the OUR at 0.15 kg(O_2)/m^3. The Reactor Temperature is at 50°C.**

The variation in the minimum influent sludge total solids concentration required to prevent the aerobic reactor from becoming substrate limited at different aerobic reactor retention times is shown graphically in Figure 5.3 below.

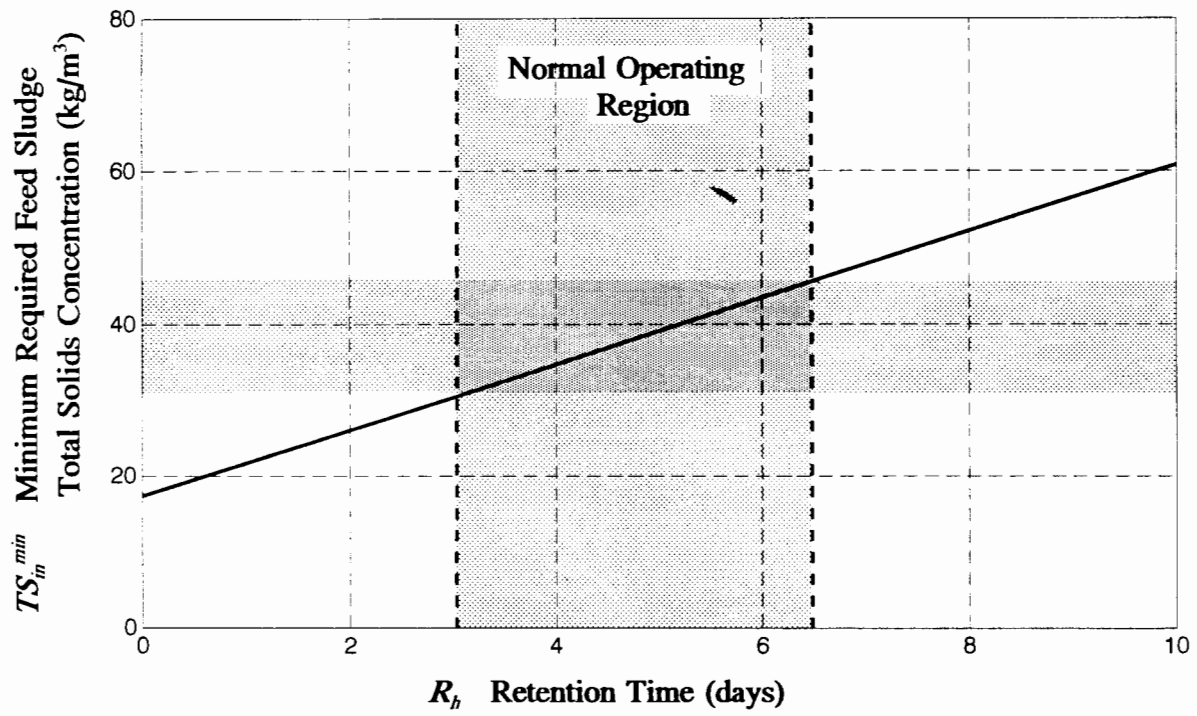

Figure 5.3 **The Variation in the Minimum Influent Sludge Total Solids Concentration required to prevent the Aerobic Reactor from becoming Substrate Limited at Different Retention Times: Oxygenation is with Air (Flow Rate 760 m^3(STP)/h) and No-Foam is Present.**

5.2.4 Volatile Solids Removal in the Aerobic Reactor

The predicted degree of volatile solids destruction which takes place in the aerobic reactor (expressed as a fraction of the influent VS) when oxygenation is with air alone follows from the above part two of model (1). Use is again made of the VS destruction

kinetic equations derived in Section 3.11. For comparison purposes, the degree of VS destruction is predicted for both oxygen and substrate limiting conditions.

VS Removal Under Oxygen Limiting Conditions

As has been stated previously, the aerobic reactor was operated under oxygen limiting conditions for each of the steady state periods covered during phase I. Accepting that the rate of volatile solids destruction in the aerobic reactor is proportional to the oxygen utilisation rate, and following on from the fact that under oxygen limiting conditions (substrate excess), the oxygen utilisation rate ***OUR*** is fixed by the maximum oxygen transfer rate OTR_{max} which the aeration device can deliver, the fraction of volatile solids removed in the aerobic reactor under oxygen limiting conditions can be expressed as a function (Eq 3.214) of the ***OUR*** and the influent sludge total solids concentration. viz:

$$f(VS)_{rem}^{O2\,lim} = \frac{24.R_h.OUR}{f_{ovs}.f_{vs}.TS_{in}} \qquad \ldots (3.214)$$

where:

$f(VS)_{rem}^{O2lim}$	= Fraction of volatile solids removed in the aerobic reactor	$f(VS)_{rem}^{O2lim}$	-
R_h	= Sludge retention time in the aerobic reactor	R_h	d
OUR	= Oxygen utilisation rate	0.15	$kg(O_2)/m^3.h$
f_{ovs}	= Mass of VS dest. per mass of oxygen utilised in the reactor	1.70	$kg(VS)/kg(O_2)$
f_{vs}	= VS fraction of the TS in the influent sludge	0.81	kg(VS)/kg(TS)

$$f(VS)_{rem}^{O2\,lim} = \frac{24.R_h \times 0.15}{1.70 \times 0.81 \times TS_{in}} \qquad \ldots (5.18)$$

which simplifies to:

$$f(VS)_{rem}^{O2\,lim} = \frac{3.6R_h}{1.38TS_{in}}$$

Equation 5.19 **The Fraction of VS Removed Under Oxygen Limiting Conditions as a function of Retention Time (d) and Influent Sludge Solids Concentration (g/l): The Reactor is Not-Foaming, the Air Flow Rate is at Maximum (760 m^3(STP)/h) and the *OUR* is fixed by the OTR_{max} at 0.15 $kg(O_2)/m^3.h$.**

For different influent sludge solids (TS_{in}) concentrations, Equation 5.19 above plots as a family of straight lines passing through the origin on the %VS removal ($100.f(VS)_{rem}^{O2lim}$) versus retention time (R_h) graph (see Figure 5.4). In Figure 5.4, 25% VS reduction will be achieved at retention times of 2.9, 3.8, and 4.8 days at influent sludge TS concentrations of 30, 40 and 50 kg/m^3 respectively (points A, B and C). Alternatively at 4 days retention, the %VS removal for 50 and 40 kg/m^3 influent sludge TS

concentrations will be 21% and 26% respectively (points D and E). At a feed sludge TS concentration of 30 kg/m^3 and a 4 day retention time the reactor is substrate limited; the curved line (Eq 3.220) for substrate limitation (point F) is below the straight line (dotted, Eq 3.214) for oxygen limitation (point G) at R_h = 4 days and TS_{in} = 30 kg/m^3. The %VS removal, calculated from VS degradation kinetics, is 30% (point F) (see below) and not 35% (point G) as predicted by Eq 5.18 which is valid only for oxygen limiting conditions. Consequently in Figure 5.4, all points (solid straight lines) below the curved (substrate limitation) line represent oxygen limitation conditions; operation above the curved line (dashed lines) is not possible because substrate limiting conditions constrain operation onto the curved line. The curved substrate limitation line therefore bounds the operation of the aerobic reactor i.e. operation is either on it, in which case substrate limitation (and oxygen sufficiency) conditions prevail, or below the line, in which case oxygen limitation (and substrate sufficiency) conditions prevail.

VS Removal Under Substrate Limiting Conditions

The fraction of volatile solids destroyed in the aerobic reactor under substrate limiting conditions is predicted by employing the VS degradation kinetics (see Section 3.11.4) of Andrews and Kambhu (1971), developed to model the ATAD process which operates under substrate limiting conditions viz:

$$f(VS)_{rem}^{VS\,lim} = \frac{k_d . f_{bio} . R_h}{(1 + R_h . k_d)} \qquad \text{... (3.220)}$$

where:

$f(VS)_{rem}^{VS\,lim}$	= Fraction of volatile solids removed in the aerobic reactor.	$f(VS)_{rem}^{O2lim}$	-
f_{bio}	= Biodegradable fraction of the VS in the influent sludge	0.60	kg(BVS)/kg(VS)
R_h	= Sludge retention time through the aerobic reactor	R_h	d
k_d	= **Volatile solids destruction rate coefficient at 50°C**	**0.25**	**d^{-1}**

$$f(VS)_{rem}^{VS\,lim} = \frac{0.60 \times R_h \times 0.25}{1 + R_h \times 0.25} \qquad \text{...MJ/h (5.20)}$$

which simplifies to:

$$f(VS)_{rem}^{VS\,lim} = \frac{0.15 R_h}{1 + 0.25 R_h}$$

Equation 5.21 **The Fraction of Volatile Solids Removed Under Substrate Limiting Conditions as a function of the Retention Time (d): No-Foam is present in the reactor, the Air Flow Rate is at maximum (760 m^3 (STP)/h) and the *OUR* is less than the OTR_{max} of 0.15 kg(O_2)/m^3.h**

The variation in the degree of volatile solids destruction in the aerobic reactor as a function of the retention time is shown graphically in Figure 5.4 below and plots as a parabolic curve starting at the origin and progressively flattening as retention time increases. In Figure 5.4, the points at which the oxygen limitation (straight) lines and substrate limitation (curved) line intersect are retention times at which the oxygen and substrate limitation conditions are equal for the different feed sludge concentrations.

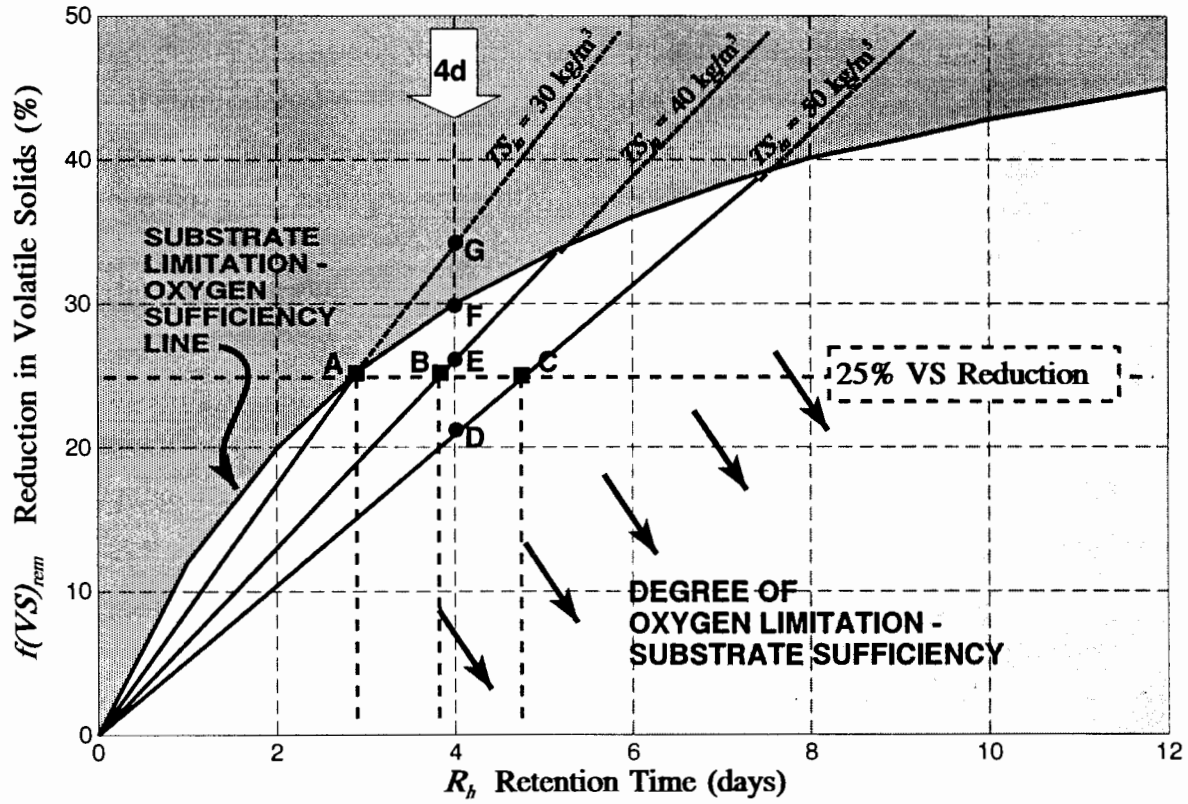

Figure 5.4 **The Variation in Volatile Solids Removal for both Oxygen and Substrate Limiting Conditions at different Aerobic Reactor Retention Times. Oxygenation is with Air (Flow Rate 760 m³ (STP)/h and No-Foam is Present.**

Figure 5.4 complements Figure 5.3 where the minimum feed sludge concentration to avoid substrate limiting conditions was plotted versus retention time (Section 5.2.3). It is interesting to note that under substrate limiting conditions the degree of VS removal is independent of the influent feed sludge solids concentration i.e. at a specific retention time, a constant %VS removal is obtained irrespective of the influent BVS concentration; the higher this concentration the greater the mass of BVS removal at constant retention time.

5.2.5 The Increase in Ammonium Ion Concentration after Aerobic Treatment

The destruction of volatile solids in the aerobic reactor gives rise to a release of ammonia into the sludge liquid (described in Section 3.1.9). Consequently the ammonia level in the sludge increases as a result of treatment in the aerobic reactor (this is one of the major conditioning effects of aerobic treatment; refer Section 3.1.9). The increase in ammonia concentration under oxygen limiting conditions can be calculated by simply extending the kinetics derived for predicting the rate of volatile solids destruction under oxygen limiting conditions (see Section 3.11.6) viz.

$$\Delta_{NH4} = [NH_4^+]_{out} - [NH_4^+]_{in} = \frac{24 . f_n . R_h . OUR}{f_{ovs}} \quad \text{...g(NH}_4\text{-N)/m}^3 \quad (3.229)$$

where:

Δ_{NH4}	= Increase in ammonia concentration after aerobic pre-treatment.	Δ_{NH4}		g(NH_4^+- N)/m^3
f_{ovs}	= Mass of VS dest. per mass of oxygen utilised in the reactor	1.70.		kg(VS)/kg(O_2)
f_n	= Nitrogen fraction of the VS in the influent primary sludge	30		g(NH_4^+- N)/kg(VS)
OUR	= Oxygen utilisation rate	0.15		kg(O_2)/m^3.h.
R_h	= Hydraulic retention time	R_h		d

$$\Delta_{NH4} = \frac{24 \times 50 \times 0.15 \times R_h}{1.70} \quad \text{...g(NH}_4\text{-N)/m}^3 \quad (5.22)$$

which simplifies to:

$$\Delta_{NH4} = 106 R_h$$

Equation 5.23 **The Increase in the Ammonium Ion Concentration (g(NH_4^+ - N)/m^3) in the Sludge after Aerobic Treatment as a Function of the Retention Time (d): Valid for Non-Foaming conditions with the Air Flow Rate at maximum (760 m^3(STP)/h) and with the *OUR* constant at 0.15 kg(O_2)/m^3.h.**

The change in the increase in ammonium concentration as a function of the retention time is shown graphically in Figure 5.5 below.

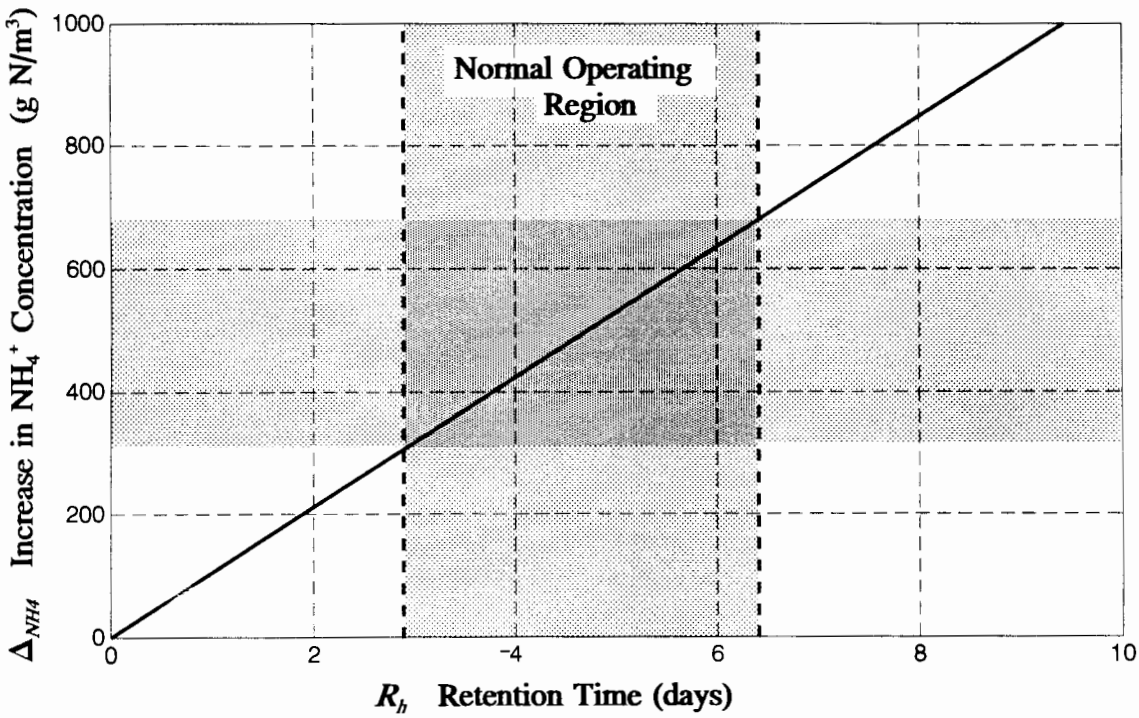

Figure 5.5 The Variation in the Increase in Ammonium Ion Concentration in the Sludge after Aerobic Treatment at different Aerobic Reactor Retention Times. Oxygenation is with Air (Flow Rate 760 m^3(STP)/h) and no Foam is Present.

5.3 MODELLING FOAMING CONDITIONS

5.3.1 Observations Made During Phase I of the Investigation

During phase I it was found that when the aerobic reactor temperature rose above 50°C the reactor started foaming. To prevent spillage, the air flow rate to the reactor was reduced which reduced the oxygen supply rate *(OSR)*. Inspite of reducing the *OSR* the oxygen utilisation rate *OUR* remained relatively unchanged compared with the *OUR* measured under non-foaming conditions. Due to the lower air flow rate through the reactor the heat losses from the reactor were accordingly lower, thus higher reactor temperatures were observed during periods of foaming.

Due to the unpredictable nature of the foaming phenomenon, plus the fact that insufficient data was obtained during the evaluation period (foaming only occurred during two of the steady state periods) it was considered unwise to attempt to make

accurate predictions regarding the operation of the aerobic reactor under foaming conditions. Consequently, two situations are presented below to give a simple illustration of the benefits of foaming and to give an indication of the magnitude of those benefits. The second model, also in two parts, predicts for varying oxygen supply rates:

- The effect of foaming on **aerobic reactor temperature** for a specific loading rate and ambient temperature.

- The effect on the **maximum loading rate** which can maintain the reactor temperature at 50°C at a specific ambient temperature.

In modelling the performance of the aerobic reactor under non-foaming conditions (Section 5.2), it was accepted that the air flow rate was fixed at 760 m^3(STP)/h giving an ***OUR*** of 0.15kgO/m^3.h. In modelling the performance of the aerobic reactor under foaming conditions, the position becomes more complicated because the aeration characteristics of the system do not remain constant. As discussed, to prevent foam spillage, the ***OSR*** is reduced (through reducing the air flow rate) and there is a change (increase) in the oxygen transfer efficiency ***(OTE)***.

To enable the reactor to be modelled under foaming conditions (and to make comparison with non-foaming conditions) simple exponential functions are derived to describe the aeration characteristics of the system, i.e. the ***OUR*** ***(=OTR)*** and ***OTE*** as functions of ***OSR*** in the form (see Figure 5.6):

$$OTE = a \,.\, \exp(b,OSR) \qquad OUR = OSR \,.\, c \,.\, \exp(\, b,OSR) \qquad \text{...viz:}$$

Aeration Characteristics: Foaming Conditions

$$OTE_{FOAM} = 43.0 \times e^{-0.822\,OSR} \qquad \text{...\%} \quad (5.24)$$

$$OUR_{FOAM} = OSR \times \frac{OTE_{FOAM}}{100} = OSR \times 0.430 \times e^{-0.822\,OSR} \qquad \text{...kg}(O_2)/m^3.h \quad (5.25)$$

Aeration Characteristics: Non-Foaming Conditions

$$OTE_{NO\text{-}FOAM} = 18.1 \times e^{-0.321\,OSR} \qquad \text{...\%} \quad (5.26)$$

$$OUR_{NO\text{-}FOAM} = OSR \times \frac{OTE_{NO\text{-}FOAM}}{100} = OSR \times 0.181\, e^{-0.321\,OSR} \qquad \text{...kg}(O_2)/m^3.h \quad (5.27)$$

The above functions were derived from the averages of the aeration characteristics *OSR*, *OTE* and *OUR* observed during the 8 steady state periods of phase I (see Table 4.3). The best fitted functions and the experimental data for the foaming and non-foaming conditions are shown in Figure 5.6 below. The correlation coefficient (R^2) for the non-foaming condition was 0.96 and for the foaming condition was 1.00 because there are only two points.

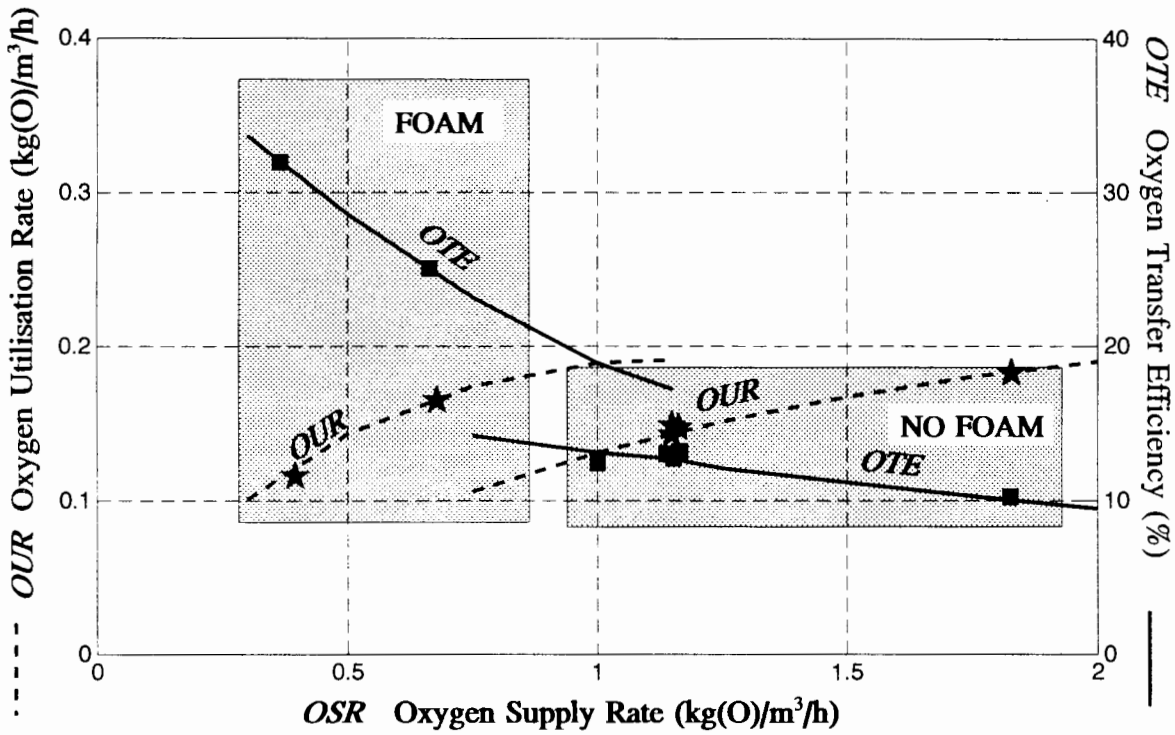

Figure 5.6 The Variation in Oxygen Utilisation Rate *(OUR)* and Oxygen Transfer Efficiency *(OTE)* at Different Oxygen Supply Rates *(OSR)* for both Foaming and Non-Foaming Conditions (data from Table 4.3).

The oxygen supply rate *(OSR)* as a function of the influent air flow rate $Q(AIR)_{in}$ was derived in Section 3.2 and is given by:

$$OSR = 0.00151 \times Q(AIR)_{in} \quad \text{...kg}(O_2)/\text{m}^3\text{.h} \quad (3.47)$$

The oxygen utilisation rates OUR_{FOAM} and $OUR_{NO\text{-}FOAM}$ in terms of the influent air flow rate $Q(AIR)_{in}$ are therefore as follows:

$$OUR_{FOAM} = 6.49 \times 10^{-4} \times Q(AIR)_{in} \times e^{-0.00124\,Q(AIR)_{in}} = f(Q(AIR)_{in})_{FOAM}$$

...$kg(O_2)/m^3.h$ (5.28)

$$OUR_{NO\text{-}FOAM} = 2.73 \times 10^{-4} \times Q(AIR)_{in} \times e^{-0.000484\,Q(AIR)_{in}} = f(Q(AIR)_{in})_{NO\text{-}FOAM}$$

...$kg(O_2)/m^3.h$ (5.29)

The ambient temperature chosen for modelling the reactor under foaming conditions is 20°C. The saturation humidity of the influent air stream U_{in} at 20°C is calculated to be 0.0090($kg(H_2O)/kg(AIR)$.

Under foaming conditions, it is accepted that, because of the presence of foam in the reactor head space, the air stream does not undergo cooling in the reactor headspace, and consequently the effluent air stream temperature $T(AIR)_{out}$ is the same as the temperature of the sludge in the reactor $T(SL)_r$. In contrast, under non-foaming conditions, the air stream is cooled in the reactor head space and the effluent air stream temperature exits at approximately 3°C lower than the temperature of the sludge in the reactor. For ease of calculation, the effluent air stream humidity is accepted to vary linearly with the temperature of the effluent air stream where the linear function (Eq 5.30) was derived from theoretically calculated saturation humidities over the temperature range 45° to 60°C.

$$U_{in} = 0.0060\,T(SL)_{out} - 0.213 \quad \text{...}kg(H_2O)/kg(AIR) \quad (5.30)$$

Figure 5.7 below shows the variation of the effluent air stream humidity with the effluent air stream temperature (the air stream is presumed fully saturated with water vapour) and compares the linear approximation Eq 5.30 (dotted line) with the theoretically calculated humidity Eq 3.124 (solid line).

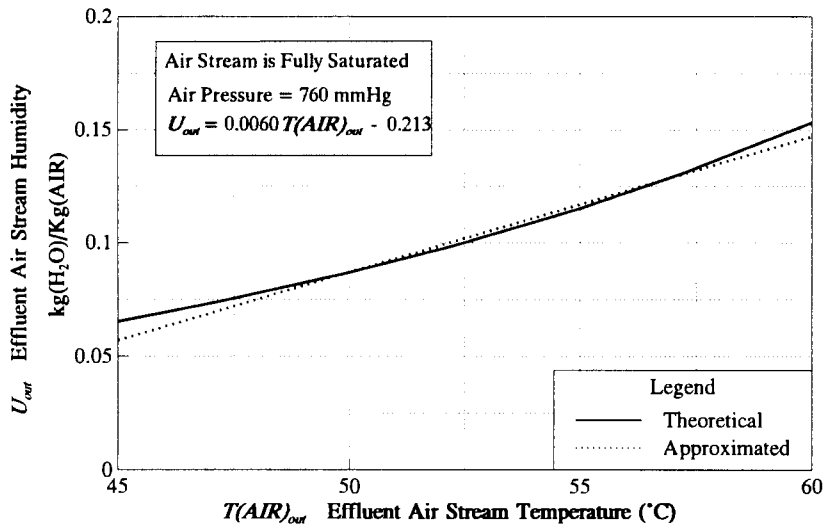

Figure 5.7 The Variation of Theoretical and Approximated Effluent Gas Stream Humidities (kg(H_2O)/kg(AIR)) with Air Stream Temperature.

5.3.2 The Effect of Foaming on Reactor Temperature

The first part of model two on foaming conditions predicts the increase in reactor temperature as the oxygen supply rate is reduced to prevent foam spillage. As discussed, during phase I, it was found that when the aerobic reactor started to foam it was necessary to reduce the air flow rate to prevent foam from spilling from the top of the reactor (see Section 4.3.5). Inspite of the reduction in the oxygen supply rate it was found that the reactor temperature actually increased as a result of the improvement in the oxygen transfer efficiency under foaming conditions. For comparison purposes, the change in reactor temperature as a result of reducing the oxygen supply rate under non-foaming conditions is also presented. The functions employed to describe the oxygen utilisation rate at different oxygen supply rates for both foaming and non-foaming conditions are as described in Section 5.3.1 above. The ambient temperature is accepted to be at 20°C with a sludge flow rate of 40 m^3/d. This loading rate is the maximum sludge flow which will allow the reactor temperature to reach 50°C at an ambient temperature of 20°C during non-foaming conditions (see Table 5.2). The prediction is based on the solution of the Steady State Heat Balance arranged in terms of the temperature of the reactor sludge $T(SL)_r$. In the derivation each heat term in the balance is described separately. It is assumed that there is no substrate limitation.

The Biological Heating Rate

$$H_b = Y_h . V_p . OUR \quad \text{...MJ/h} \quad (5.31)$$

where:

Y_h	= Specific Heat Yield	12.8	$MJ/kg(O_2)$
V_p	= Aerobic reactor process volume	184	m^3
OUR_{FOAM}	= Oxygen utilisation rate under foaming conditions	$f(Q(AIR)_{in})_{FOAM}$	$kg(O_2)/m^3.h.$
$OUR_{NO\text{-}FOAM}$	=Oxygen utilisation rate under non-foaming conditions	$f(Q(AIR)_{in})_{NO\text{-}FOAM}$	$kg(O_2)/m^3.h.$

$$H_{b_{FOAM}} = 12.8 \times 184 \times f(Q(AIR)_{in})_{FOAM} = 1.53 \times Q(AIR)_{in} \times e^{-0.00124 Q(AIR)_{in}} \quad \text{...MJ/h} \quad (5.32)$$

$$H_{b_{NOFOAM}} = 12.8 \times 184 \times f(Q(AIR)_{in})_{NOFOAM} = 0.642 \times Q(AIR)_{in} \times e^{-0.000485 Q(AIR)_{in}} \quad \text{...MJ/h} \quad (5.33)$$

The Mechanical Heat Input Rate

$$H_m = 0.0036 \times f_{mech} . \cos\phi . \sqrt{3} . V . I \quad \text{...MJ/h} \quad (5.34)$$

where:

f_{mech}	= Fraction of electrical energy converted to heat	0.85	
V	= Phase voltage of the pump motor	380	volts
I	= Phase current drawn by the pump motor	19	amps
$cos\Phi$	= Power factor for the pump motor	0.82	
0.0036	= Conversion factor from W to MJ/h	0.0036	MJ/W.h

$$H_m = 0.0036 \times 0.85 \times 0.82 \times 1.732 \times 380 \times 19 = 31 \quad \text{...MJ/h} \quad (5.35)$$

The Effluent Gas Vapour Heat Loss Rate

$$H_v = L_v^{50°C} . \rho(AIR) . Q(AIR)_{in} . (U_{out} - U_{in}) \quad \text{...MJ/h} \quad (5.36)$$

where:

$L_v^{50°C}$	= Latent heat of vaporisation of water at 50°C	2.38	$MJ/kg(H_2O)$
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	$kg(AIR)/m^3$
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	760	$m^3(STP)/h$
$U_{outFOAM}$	= Effluent air stream humidity under foaming conditions	$f(T_r)$	$kg(H_2O)/kg(AIR)$
$U_{outNO\text{-}FOAM}$	=Effluent air stream humidity under non-foaming conditions	$f(T_r\text{-}3)$	$kg(H_2O)/kg(AIR)$
U_{in}	= Influent air stream humidity (function of T_{amb})	$f(T_{amb})$	$kg(H_2O)/kg(AIR)$

$$H_{v_{FOAM}} = 2.38 \times 1.205 \times Q(AIR)_{in} ((0.0060 T_r - 0.213) - 0.0090) = Q(AIR)_{in} (0.0172 T_r - 0.637) \quad \text{...MJ/h} \quad (5.37)$$

$$H_{v_{NO\ FOAM}} = 2.38 \times 1.205 \times Q(AIR)_{in} ((0.0060 (T_r - 3) - 0.213) - 0.0090) = Q(AIR)_{in} (0.0172 T_r - 0.688) \quad \text{...MJ/h} \quad (5.38)$$

The Effluent Gas Sensible Heat Loss Rate

$$H_g = \rho(AIR).Q(AIR)_{in}.\left(C_p^{20^\circ C}(AIR).(T(AIR)_r - T(AIR)_{in}) + U_{out}.C_p^{50^\circ C}(H_2O)_{vap}.T(AIR)_r\right) \text{...MJ/h} \quad (5.39)$$

where:

ρ *(AIR)*	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	kg(AIR)/m^3
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	760	m^3(STP)/h
$C_p^{20^\circ C}$ (AIR)	= Heat capacity of dry air at STP	0.0010	MJ/kg(AIR)
$C_p^{50^\circ C}$ $(H_2O)_{vap}$	=Heat capacity of water vapour at 50°C	0.00187	MJ/kg(H_2O)
$U_{outFOAM}$	= Effluent air stream humidity under foaming conditions	$f(T_r)$	kg(H_2O)/kg(AIR)
$U_{outNO\text{-}FOAM}$	=Effluent air stream humidity under non-foaming conditions	$f(T_r$-3)	kg(H_2O)/kg(AIR)
$T(AIR)_{in}$	= Influent air stream temperature	20	°C
$T(AIR)_r$	= Effluent air stream temperature	T_r	°C

$$H_{g_{FOAM}} = 1.205 \times Q(AIR)_{in}\left(0.0010(T_r - 20) + (0.006T_r - 0.213) \times 0.00187 \times T_r\right)$$

$$= Q(AIR)_{in}\left(1.35 \times 10^{-5}T_r^2 + 7.254 \times 10^{-4}T_r - 0.0241\right) \text{...MJ/h} \quad (5.40)$$

$$H_{g_{NO\text{-}FOAM}} = 1.205 \times Q(AIR)_{in}\left(0.001((T_r - 3) - 20) + (0.006(T_r - 3) - 0.213) \times 0.00187\right.$$

$$= Q(AIR)_{in}\left(1.35 \times 10^{-5}T_r^2 + 6.5808 \times 10^{-4}T_r - 0.0277\right) \text{...MJ/h} \quad (5.41)$$

The above expressions (Eq's 5.40 and 5.41) can be approximated by the following linear functions, valid between the range T_r = 45°C to 60°C with the maximum error less than 1 MJ/h:

$$H_{g_{FOAM}} = Q(AIR)_{in}(0.0023T_r - 0.070) \text{...MJ/h} \quad (5.42)$$

$$H_{g_{NO\text{-}FOAM}} = Q(AIR)_{in}(0.0023T_r - 0.077) \text{...MJ/h} \quad (5.43)$$

The Effluent Sludge Sensible Heat Loss Rate

$$H_s = C_p^{20^\circ C}(SL)\left(\left(\frac{Q(SL)_{in}}{24} - \frac{(\rho(AIR).Q(AIR)_{in}.U_{out}}{1000}\right).T(SL)_r - \frac{Q(SL)_{in}.T(SL)_{in}}{24}\right) \text{...MJ/h} \quad (5.44)$$

where:

$C_p^{20^\circ C}(SL)$	= Heat capacity of sludge at 20°C	4.0	MJ/m^3
$Q(SL)_{in}$	= Influent sludge stream flow rate	Q_i^{max}	m^3/d
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.205	kg(AIR)/m^3
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	760	m^3(STP)/h
$U_{outFOAM}$	= Effluent air stream humidity under foaming conditions	$f(T_r)$	kg(H_2O)/kg(AIR)
$U_{outNO\text{-}FOAM}$	= Effluent air stream humidity under non-foaming conditions	$f(T_r$-3)	kg(H_2O)/kg(AIR)
$T(SL)_{in}$	= Influent sludge stream temperature	20	°C
$T(SL)_r$	= Effluent sludge stream temperature	T_r	°C

$$H_{s_{FOAM}} = 4.0\left(\left(\frac{40}{24} - \frac{1.205 \times Q(AIR)_{in} \times (0.0060T_r - 0.213)}{1000}\right)T_r - \frac{40 \times 20}{24}\right)$$

$$= 6.667T_r - 2.89 \times 10^{-5}Q(AIR)_{in}T_r^2 + 0.00103Q(AIR)_{in}T_r - 133.1 \text{...MJ/h} \quad (5.45)$$

$$H_{s_{NO\ FOAM}} = 4.0\left(\left(\frac{40}{24} - \frac{1.205 \times Q(AIR)_{in} \times (0.0060(T_r - 3) - 0.213)}{1000}\right)T_r - \frac{40 \times 20}{24}\right)$$

$$= 6.667T_r - 2.89 \times 10^{-5}Q(AIR)_{in}T_r^2 + 0.00112Q(AIR)_{in}T_r - 133.1 \text{...MJ/h} \quad (5.46)$$

The above expressions (Eq's 5.45 and 5.46) can be approximated by the following linear functions, valid between the range T_r = 45°C to 60°C with the maximum error less than 1 MJ/h:

$$H_{s_{FOAM}} = 6.667T_r - 0.00234Q(AIR)_{in}T_r + 0.0971Q(AIR)_{in} - 133.1 \quad \text{...MJ/h} \quad (5.47)$$

$$H_{s_{NO\text{-}FOAM}} = 6.667T_r - 0.00225Q(AIR)_{in}T_r + 0.0971Q(AIR)_{in} - 133.1 \quad \text{...MJ/h} \quad (5.48)$$

The Wall Heat Loss Rate

$$H_w = 0.82(T(SL)_r - T(SL)_d) + 0.49(T(SL)_r - T(AIR)_{in}) \quad \text{...MJ/h} \quad (5.49)$$

where:

$T(SL)_r$	= Aerobic reactor sludge temperature	T_r	°C
$T(AIR)_{in}$	= Ambient air temperature	20	°C
$T(SL)_d$	= Anaerobic digester sludge temperature	$(T_r+20)/2$	°C

$$H_w = 0.82(T_r - (T_r + 20)/2) + 0.49(T_r - 20) = 0.9T_r - 18 \quad \text{...MJ/h} \quad (5.50)$$

Combining each of the above heat terms in the steady state heat balance, firstly for foaming conditions and there for non-foaming conditions yields:

Heat Balance for Foaming Conditions

$$H_{b_{FOAM}} + H_m = H_{v_{FOAM}} + H_{g_{FOAM}} + H_{s_{FOAM}} + H_w \quad \text{...MJ/h} \quad (3.8)$$

$$1.53 \times Q(AIR)_{in} \times e^{-0.00124Q(AIR)_{in}} + 31 = Q(AIR)_{in}(0.0172T_r - 0.637) + Q(AIR)_{in}(0.0023T_r - 0.070)$$
$$+ 6.667T_r - 0.00234Q(AIR)_{in}T_r + 0.0971Q(AIR)_{in} - 133.1 + 0.9T_r - 18 \quad \text{...MJ/h} \quad (5.51)$$

Rearranging in terms of the aerobic reactor temperature T_r yields:

$$T_r = \frac{1.53 \times Q(AIR)_{in} \times e^{-0.00124Q(AIR)_{in}} + 182 + 0.610Q(AIR)_{in}}{7.57 + 0.0172Q(AIR)_{in}}$$

Equation 5.52 **The Aerobic Reactor Temperature (T_r) as a Function of the Influent Air Stream Flow Rate $Q(AIR)_{in}$ (m^3(STP)/h) under Foaming Conditions with an Ambient Temperature of 20°C and a Sludge Loading Rate of 40m^3/d of Feed Sludge.**

Heat Balance for Non-Foaming Conditions

$$H_{b_{NO\text{-}FOAM}} + H_m = H_{v_{NO\text{-}FOAM}} + H_{g_{NO\text{-}FOAM}} + H_{s_{NO\text{-}FOAM}} + H_w \quad \text{...MJ/h} \quad (3.8)$$

$$0.642 \times Q(AIR)_{in} \times e^{-0.000485\,Q(AIR)_{in}} + 31 = Q(AIR)_{in}(0.0172T_r - 0.688) + Q(AIR)_{in}(0.0023T_r - 0.077)$$

$$+ 6.667T_r - 0.00225\,Q(AIR)_{in}T_r + 0.0971\,Q(AIR)_{in} - 133.1 + 0.9T_r - 18 \quad \text{...MJ/h} \quad (5.54)$$

Rearranging in terms of the aerobic reactor temperature T_r yields:

$$T_r = \frac{0.642 \times Q(AIR)_{in} \times e^{-0.000485\,Q(AIR)_{in}} + 182 + 0.668\,Q(AIR)_{in}}{7.57 + 0.0172\,Q(AIR)_{in}}$$

Equation 5.55 **The Aerobic Reactor Temperature (T_r) as a Function of the Influent Air Stream Flow Rate $Q(AIR)_{in}$ (m^3(STP)/h) under Non-Foaming Conditions with an Ambient Temperature of 20°C and a Sludge Loading Rate of 40m^3/d of Feed Sludge (see Eq 5.52).**

A graphical illustration of the variation in aerobic reactor temperature (T_r) at different influent air flow rates m^3(STP)/h under both foaming and non-foaming conditions with an ambient temperature of 20°C is shown in Figure 5.8 below. In this example the sludge feed flow rate was taken as 40 m^3/d (R_h = 4.6d) which at the ambient temperature selected of 20°C and an air flow rate of 760 m^3(STP)/h under non-foaming conditions is predicted to produce a reactor temperature of 50°C (see Table 5.2) (point A on Figure 5.8).

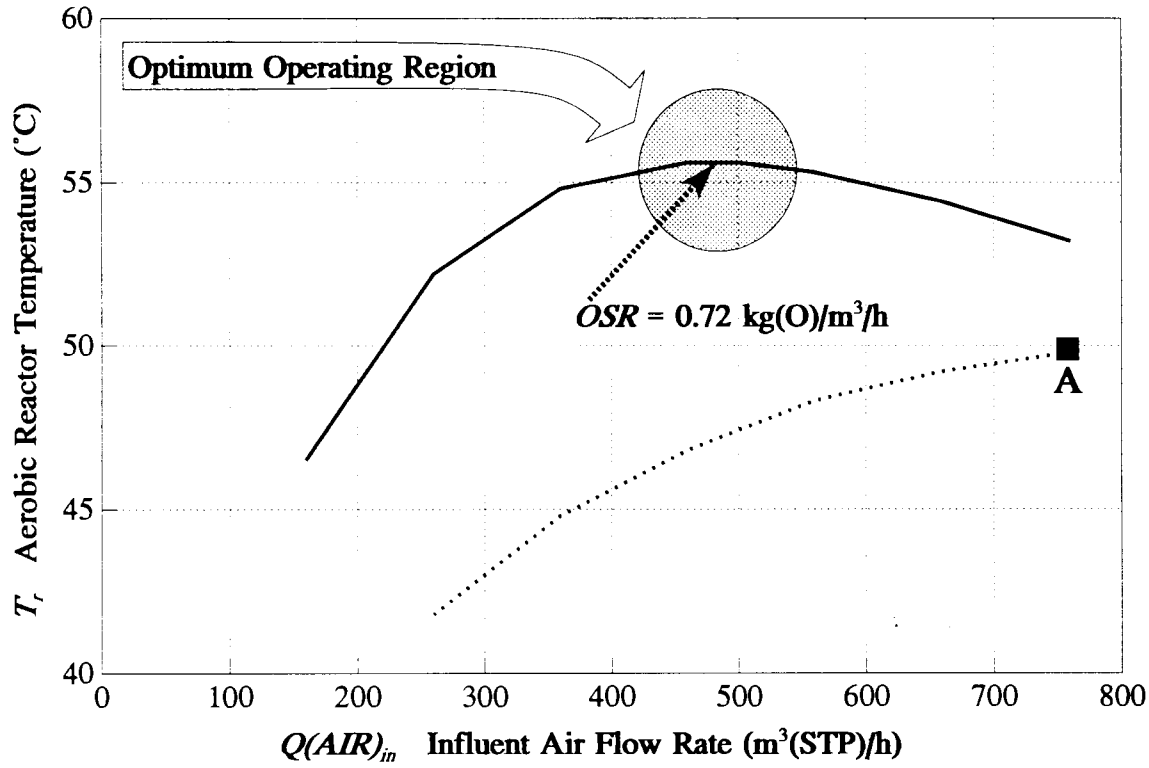

Figure 5.8 **The Variation in Aerobic Reactor Temperature (T_r) at differing influent air stream flow rates m^3(STP)/h under Foaming and Non-Foaming Conditions with an Ambient Temperature of 20°C and a Loading Rate of 40m^3/d of Feed Sludge.**

From Figure 5.8, an optimum air flow rate ($Q(AIR)_{in}^{opt}$) of 480 m^3(STP)/h is predicted which is equivalent to an *OSR* = 0.725 kg(O_2)/m^3.h and slightly higher than the average *OSR* observed during steady state period 4 when foaming occurred (and a reactor temperature of 54.2°C was measured; see Tables 4.3 and 4.8 and Figure 5.6). At this air flow rate maximum benefit will be derived from the foaming phenomenon in terms of reactor temperature. The aeration characteristics at $Q(AIR)_{in}^{opt}$, calculated by Equations Eq 3.41(*OSR*), Eq 5.23(*OTE*) and Eq 5.27(*OUR*) are presented in Table 5.3 below:

Table 5.3 The Aeration Characteristics at the Optimum Air Flow Rate ($Q(AIR)_{in}^{opt}$) of 480 m^3(STP)/h for Operation under Foaming Conditions.

Optimum Air Flow Rate: $Q(Air)_{in}^{opt}$ m^3(STP)/h	Oxygen Supply Rate : *OSR* $kg(0_2)/m^3.h$	Oxygen Transfer Efficiency: *OTE* %	Oxygen Utilisation Rate: *OUR* $kg(0_2)/m^3.h$
480	0.725	23.7	0.172

5.3.3 The Effect on Foaming on the Maximum Sludge Loading Rate

The second part of the (second) model on foaming predicts the increase in the maximum sludge loading rate (the loading rate required to maintain a reactor temperature of 50°C) made possible by the improved aeration characteristics brought about by the presence of foam. As with the previous model presented in Section 5.3.2 above, it is assumed that the air flow rate (i.e. the oxygen supply rate *OSR)* is reduced to prevent foam spillage. However, inspite of the reduction in the oxygen supply rate, aerobic reactor performance is enhanced and it is anticipated that increased loading should be possible if the foam layer can be maintained. Increased sludge loading is possible because the biological heating rate is not significantly affected by the reduction in the oxygen supply rate due to the increase in the oxygen transfer efficiency *OTE*. A reduced air flow rate through the system results in a lowering of the heat losses in the effluent air stream and this makes it possible to increase the sludge sensible heat loss from the system H_s (i.e. increase the sludge loading rate) without causing a drop in the reactor temperature below 50°C.

The model predicts the anticipated increase in the maximum sludge loading rate as the oxygen supply rate is reduced under foaming conditions. For comparison purposes, the change in the maximum sludge loading rate as a result of reducing the oxygen supply rate under non-foaming conditions is also presented.

The functions employed to describe the oxygen utilisation rate at different oxygen supply rates for both foaming and non-foaming conditions are as described in Section 5.3.1 above. The ambient temperature is accepted to be at 20°C and the aerobic reactor temperature is assumed to be at 50°C. The prediction is based on the solution of the Steady State Heat Balance and use is made of the Equations derived in Section 5.3.2. above. Each heat term in the heat balance is described separately. The effluent sludge sensible heat loss rate equation (Eq 5.64) is expanded because of the variation in the

sludge flow rate $Q(SL)_{in}$. Again it is assumed that the aerobic reactor is oxygen limited and not substrate limited.

It should be noted that this model should be used with caution, as an increase in the sludge loading rate may well cause the foaming to stop, which would then cause the temperature in the reactor to decrease.

The Biological Heating Rate

$$H_{b_{FOAM}} = 1.53 \times Q(AIR)_{in} \times e^{-0.00124\,Q(AIR)_{in}} \quad \text{...MJ/h} \quad (5.56)$$

$$H_{b_{NOFOAM}} = 0.642 \times Q(AIR)_{in} \times e^{-0.000485\,Q(AIR)_{in}} \quad \text{...MJ/h} \quad (5.57)$$

The Mechanical Heat Input Rate

$$H_m = 31 \quad \text{...MJ/h} \quad (5.58)$$

The Effluent Gas Vapour Heat Loss Rate

$$H_v = Q(AIR)_{in}(0.0172\,T_r - 0.637) \quad \text{...MJ/h} \quad (5.59)$$

where:

$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	$Q(AIR_{in})$	m^3(STP)/h
$T_{r\text{-}FOAM}$	= Effluent air stream temperature under foaming conditions	50	°C
$T_{r\text{-}NO\text{-}FOAM}$	=Effluent air stream temperature under non-foaming conditions	47	°C

$$H_{v_{FOAM}} = Q(AIR)_{in}(0.0172 \times 50 - 0.637) = 0.223\,Q(AIR)_{in} \quad \text{...MJ/h} \quad (5.60)$$

$$H_{v_{NO\text{-}FOAM}} = Q(AIR)_{in}(0.0172 \times 47 - 0.637) = 0.171\,Q(AIR)_{in} \quad \text{...MJ/h} \quad (5.61)$$

The Effluent Gas Sensible Heat Loss Rate

$$H_g = Q(AIR)_{in}(0.0023\,T_r - 0.070) \quad \text{...MJ/h} \quad (5.62)$$

where:

$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	$Q(AIR_{in})$	m^3(STP)/h
$T_{r\text{-}FOAM}$	= Effluent air stream temperature under foaming conditions	50	°C
$T_{r\text{-}NO\text{-}FOAM}$	=Effluent air stream temperature under non-foaming conditions	47	°C

$$H_{g_{FOAM}} = Q(AIR)_{in}(0.0023 \times 50 - 0.070) = 0.045\,Q(AIR)_{in} \quad \text{...MJ/h} \quad (5.63)$$

$$H_{g_{NO\ FOAM}} = Q(AIR)_{in}(0.0023 \times 47 - 0.070) = 0.038\,Q(AIR)_{in} \quad \text{...MJ/h} \quad (5.64)$$

The Effluent Sludge Sensible Heat Loss Rate

$$H_s = C_p^{20^\circ C}(SL)\left(\left(\frac{Q(SL)_{in}}{24} - \frac{(\rho(AIR).Q(AIR)_{in}.U_{out}}{1000}\right).T(SL)_r - \frac{Q(SL)_{in}.T(SL)_{in}}{24}\right) \quad ...MJ/h \quad (5.65)$$

where:

$C_p^{20^\circ C}(SL)$	= Heat capacity of sludge at 20°C	4.0	MJ/m^3
$Q(SL)_{in}$	= Influent sludge stream flow rate	Q_i^{max}	m^3/d
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.205	$kg(AIR)/m^3$
$Q(AIR)_{in}$	= Dry influent air stream vol. flow rate at STP	$Q(AIR)_{in}$	$m^3(STP)/h$
$U_{outFOAM}$	= Effluent air stream humidity under foaming conditions	0.087	$kg(H_2O)/kg(AIR)$
$U_{outNO\text{-}FOAM}$	= Effluent air stream humidity under non-foaming conditions	0.074	$kg(H_2O)/kg(AIR)$
$T(SL)_{in}$	= Influent sludge stream temperature	20	°C
$T(SL)_r$	= Effluent sludge stream temperature	50	°C

$$H_{s_{FOAM}} = 4.0\left(\left(\frac{Q_i^{max}}{24} - \frac{1.205 \times Q(AIR)_{in} \times 0.087}{1000}\right)50 - \frac{Q_i^{max} \times 20}{24}\right) = 5.0Q_i^{max} - 0.021Q(AIR)_{in}$$

...MJ/h (5.66)

$$H_{s_{NO\ FOAM}} = 4.0\left(\left(\frac{Q_i^{max}}{24} - \frac{1.205 \times Q(AIR)_{in} \times 0.074}{1000}\right)50 - \frac{Q_i^{max} \times 20}{24}\right) = 5.0Q_i^{max} - 0.018Q(AIR)_{in}$$

...MJ/h (5.67)

The Wall Heat Loss Rate

$$H_w = 0.9T_r - 18 \quad ...MJ/h \quad (5.68)$$

where:

T_r	= Aerobic reactor sludge temperature	50	°C

$$H_w = 0.9 \times 50 - 18 = 27 \quad ...MJ/h \quad (5.69)$$

Heat Balance for Foaming Conditions

Combining each of the above heat terms in the steady state heat balance

$$H_{b_{FOAM}} + H_m = H_{v_{FOAM}} + H_{g_{FOAM}} + H_{s_{FOAM}} + H_w \quad ...MJ/h \quad (3.8)$$

yields:

$$1.53 \times Q(AIR)_{in} \times e^{-0.00124Q(AIR)_{in}} + 31 = 0.223Q(AIR)_{in} + 0.045Q(AIR)_{in} + 5.0Q_i^{max} - 0.021Q(AIR)_{in} + 27$$

...MJ/h (5.69)

Rearranging in terms of Q_i^{max} yields:

$$Q_i^{max} = 0.306 \times Q(AIR)_{in} \times e^{-0.00124\,Q(AIR)_{in}} + 0.8 - 0.049\,Q(AIR)_{in}$$

Equation 5.70 **The Maximum Sludge Loading Rate Q_i^{max} (m^3/d) as a Function of the Influent Air Stream Flow Rate $Q(AIR)_{in}$ (m^3(STP)/h) under Foaming Conditions with an ambient temperature of 20°C. It is Presumed that the Reactor Temperature Remains Constant at 50°C.**

Heat Balance for Non-Foaming Conditions

Combining each of the above heat terms in the steady state heat balance:

$$H_{b_{NO\text{-}FOAM}} + H_m = H_{v_{NO\text{-}FOAM}} + H_{g_{NO\text{-}FOAM}} + H_{s_{NO\text{-}FOAM}} + H_w \quad \text{...MJ/h} \quad (3.8)$$

yields:

$$0.642 \times Q(AIR)_{in} \times e^{-0.000485\,Q(AIR)_{in}} + 31 = 0.171\,Q(AIR)_{in} + 0.038\,Q(AIR)_{in} + 5.0\,Q_i^{max} - 0.018\,Q(AIR)_{in} + 27 \quad \text{...MJ/h} \quad (5.71)$$

Rearranging in terms of Q_i^{max} yields:

$$Q_i^{max} = 0.128 \times Q(AIR)_{in} \times e^{-0.000485\,Q(AIR)_{in}} + 0.8 - 0.038\,Q(AIR)_{in}$$

Equation 5.72 **The Maximum Sludge Loading Rate Q_i^{max} (m^3/d) as a Function of the Influent Air Stream Flow Rate $Q(AIR)_{in}$ (m^3(STP)/h) under Non-Foaming Conditions with an ambient temperature of 20°C. It is Presumed that the Reactor Temperature Remains Constant at 50°C.**

A graphical illustration of the variation in the maximum sludge loading rate Q_i^{max} (m^3/d) to maintain the reactor temperature at 50°C for different influent air flow rates $Q(AIR)_{in}$ (m^3(STP)/h) under both foaming and non-foaming conditions, with an ambient temperature of 20°C, is shown in Figure 5.9 below. An optimum air flow rate ($Q(AIR)_{in}^{opt}$) of 560 m^3(STP)/h is predicted which is equivalent to an ***OSR*** = 0.846 kg(O_2)/m^3.h. Under non-foaming conditions the maximum sludge loading rate (Q_{in}^{max}) increases with increasing air flow rate ($Q(AIR)_{in}$) and reaches its limit when the air flow rate is at 760 m^3(STP)/h (the maximum flow rate which the compressor can deliver).

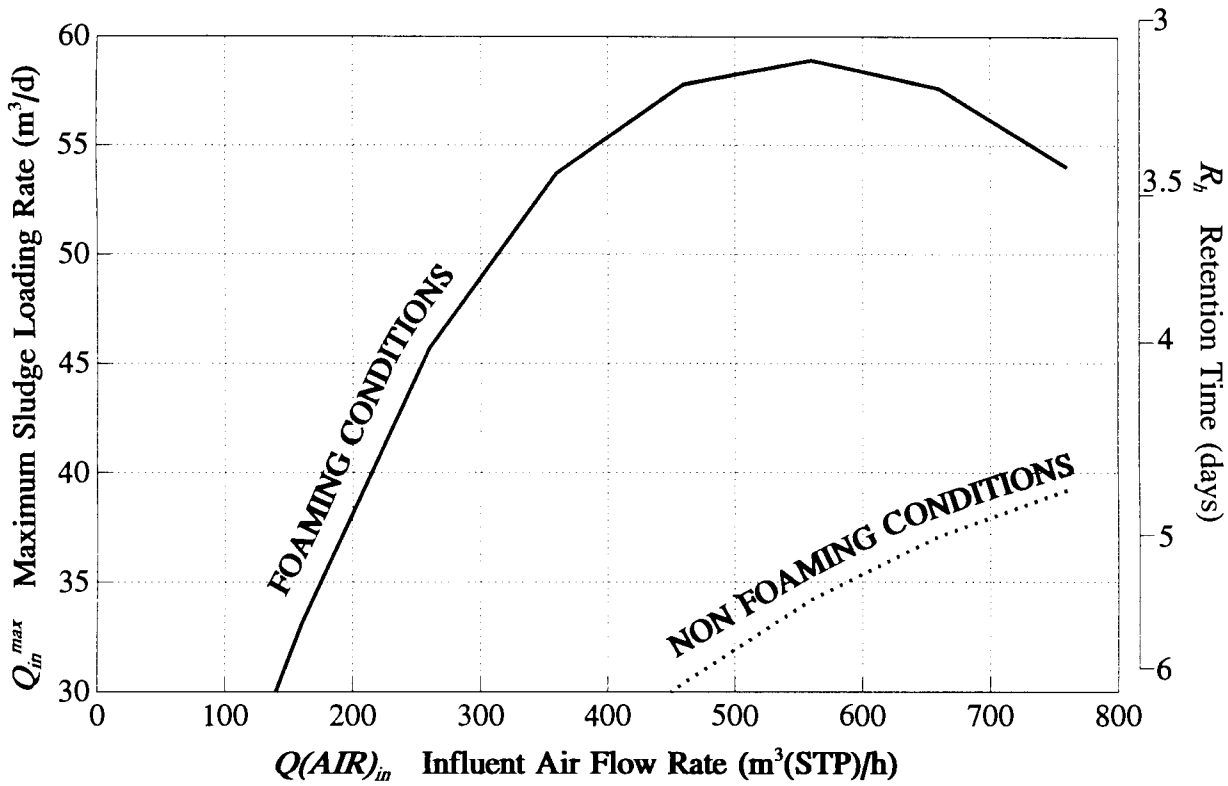

Figure 5.9 **The Variation in the Maximum Sludge Loading Rate Q_i^{max} (m³/d) at Differing Influent Air Flow Rates $Q(AIR)_{in}$ (m³(STP)/h) under Foaming and Non-Foaming Conditions with an ambient temperature of 20°C. It is Presumed that the Reactor Temperature Remains Constant at 50°C**

5.4 PURE OXYGEN SUPPLEMENTATION

5.4.1 The Motivation for Pure Oxygen Supplementation

During phase I, under non-foaming conditions, with the influent air flow rate to the aerobic reactor set at 760 m³(STP)/h, the oxygen utilisation rate ***OUR*** remained relatively constant at 0.15 kg(O_2)/m³.h. Consequently the biological heating rate H_b is fixed under such conditions at 353 MJ/h (see Eq 5.3). This in turn limits the minimum sludge retention time at which it is possible to reach disinfecting temperatures (see Section 5.2.2 above).

Further reduction in the minimum sludge retention time can be achieved only by either (1) the input of additional external heat into the system (e.g. pre-heating the feed sludge)

or by (2) increasing the biological heating rate H_b. The latter can be achieved by providing additional oxygenation with pure or semi-pure oxygen.

5.4.2 Oxygen Injection Rates to Increase Loading: Non-Foaming Conditions

To enable the pure oxygen equipment for phase II of the investigation to be properly sized, the following model[4] was developed to predict the required rates of pure oxygen injection (pure oxygen supply rates) in order to reduce the minimum sludge retention time for operation of the reactor at 50°C and 60°C, at differing ambient temperatures.

The third model is based on the solution of the steady state heat balance arranged in terms of R_h. Excluding the biological heating rate (which increases due to the utilisation of the pure oxygen), all other terms in the heat balance are accepted to remain constant (at 50° or 60°C respectively) and are therefore as described in Section 5.2.2 above; It is accepted that the masses of gaseous carbon dioxide generated from the utilisation of the pure oxygen and the remaining (un-utilised) gaseous pure oxygen are negligible (<7%) in comparison to the air stream mass flow rate through the reactor. Consequently there is no change in the gas sensible and vapour heat losses from the system with pure oxygen supplementation at a constant reactor temperature. Further it is accepted that the mechanical heat input to the system does not change.

In calculating the increase in the biological heating rate, the heat generated from (1) the utilisation of the oxygen in the air H_b^{AIR} and (2) the utilisation of the pure oxygen H_b^{O2} are treated independently. The transfer efficiency of the pure oxygen (OTE^{O2}) is assumed to be 80%. This figure is the lowest value observed by Messenger *et al* (1992) at high *OTR*'s close to oxygen sufficiency conditions. It is assumed that there is no limitation on available substrate. Whist the calculations presented below are based on maintaining a reactor temperature of 50°C, pure oxygen supply rates required to maintain the temperature at 60°C are provided for information and comparison in Table 5.4 below. Practically, for the Athlone plant to qualify as a pasteurisation system it needs to be run at 60°C.

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[4] A more precise model which predicts pure oxygen injection rates under differing operating conditions is presented in Chapter 8 below. This model is calibrated using the data obtained during phase II and employs the relevent equations developed in Chapter 3 above. Consequently, this eliminates the need to make approximations with regard to the gas flow rate and acknowledges the change in mechanical heat input, via the pumped recirculation oxygen injection system..

The Biological Heating Rate (with Oxygen Supplementation)

$$H_b = H_b^{AIR} + H_b^{O2} = Y_h.V_p.OUR^{AIR} + Y_h.OTE^{O2}.M(O_2)_{in}^{O2} \quad \text{...MJ/h} \quad (5.73)$$

where:

Y_h	= Specific heat yield	12.8	MJ/kg(O_2)
V_p	= Aerobic reactor process volume	184	m^3
OUR	= Oxygen utilisation rate (with air)	0.15	kg(O_2)/m^3.h
$M(O_2)_{in}^{O2}$	= Mass (pure) oxygen supplementation rate	$M(O_2)_{in}^{O2}$	kg(O_2)/h
OTE^{O2}	= Pure oxygen transfer efficiency	0.80	-

$$H_b = 12.8 \times 184 \times 0.15 + 12.8 \times 0.80\, M(O_2)_{in}^{O2} = 353 + 10.2\, M(O_2)_{in}^{O2} \quad \text{...MJ/h} \quad (5.74)$$

Expanding the steady state heat balance to incorporate both the biological heat terms H_b^{AIR} and H_b^{O2} yields:

$$H_b^{AIR} + H_b^{O_2} + H_m = H_v + H_g + H_s + H_w \quad \text{...MJ/h} \quad (5.75)$$

Incorporating the solution for the combined biological heating rate (Eq 5.74) into the solution of the steady state heat balance (Eq 5.14) derived in Section 5.2.2 yields:

$$353 + 10.2M(O_2)_{in}^{O2} + 31 = 166 - 1.26T_{amb} + 49 - 0.92T_{amb} + 8.33Q(SL)_{in} - 14 - 0.167Q(SL)_{in}T_{amb} + 45 - 0.9T_{amb}$$

...MJ/h (5.76)

Rearranging in terms of $M(O2)_{in}^{O2}$ yields:

$$M(O_2)_{in}^{O2} = 0.81Q(SL)_{in} - 0.30T_{amb} - 0.0163Q(SL)_{in}T_{amb} - 13.5 \quad \text{...kg(}O_2\text{)/h} \quad (5.77)$$

Which in terms of retention time R_h (note $Q(SL)_{in} = V_p / R_h = 184 / R_h$) is given by:

$$M(O_2)_{in}^{O2} = \frac{184}{R_h} \times (0.81 - 0.0163T_{amb}) - 0.30T_{amb} - 13.5$$

Equation 5.78 **The Rate of Pure Oxygen Supplementation Required (kg(O2)/h) to Achieve a Reactor Temperature of 50°C as a function of the Retention Time and Ambient Temperature under Non-Foaming Conditions.**

The variation in the required rate of oxygen supplementation, to allow the aerobic reactor to reach an operating temperature of 50°C, for specific retention times between 3 and 1 days, at different ambient temperatures is illustrated in Figure 5.10 below. Note that this graph is valid for conditions when foam is not present, and the single air compressor is operating at its maximum air flow rate of 760 m^3(STP)/h.

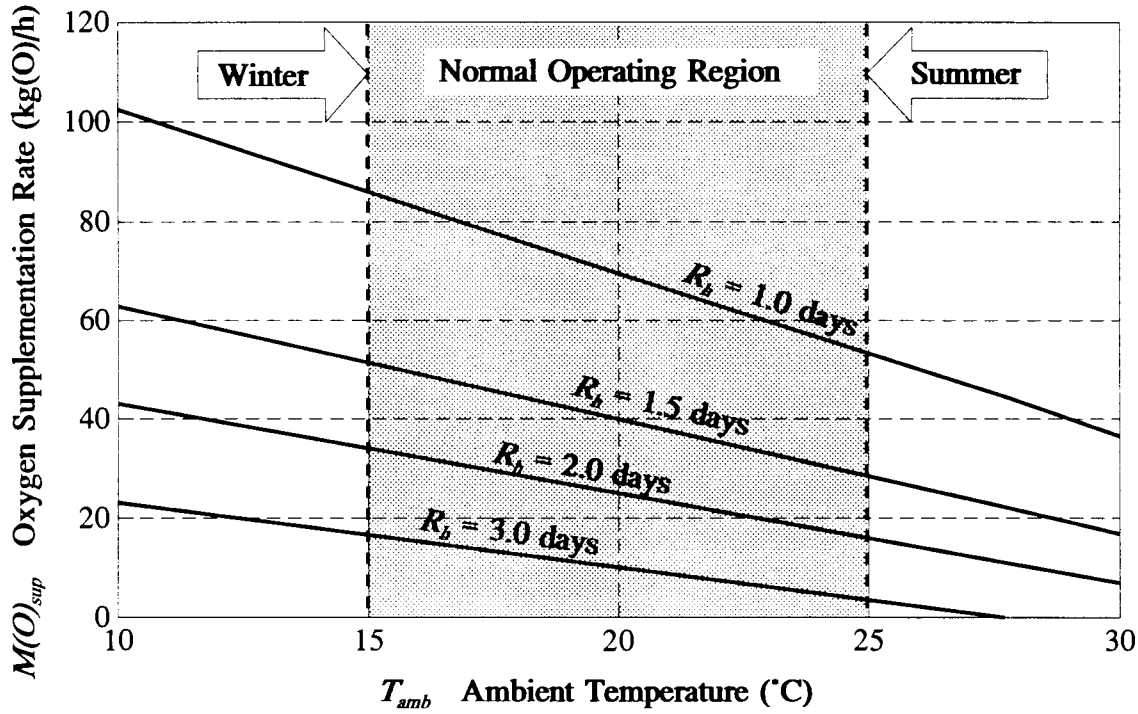

Figure 5.10 The Variation in the Required Rate of Oxygen Supplementation which will allow a Reactor Temperature of 50°C with Ambient Temperature for Different Sludge Retention Times: The Air Flow Rate is set at 760 m^3(STP)/h with No-Foam Present and No Substrate Limitation.

For design purposes, accepting an average ambient temperature of 25°C in summer and 15°C in winter, Table 5.4 below gives the required oxygen supplementation rates to maintain the aerobic reactor at both 50°C and 60°C for different sludge retention times. The data is valid for an influent air flow rate of 760 m^3(STP)/h with no foam on the surface of the reactor, and no substrate limitation.

Table 5.4 **The Required Oxygen Supplementation Rates to Maintain the Aerobic Reactor Temperature at 50°C and 60°C for the Sludge Retention Times Specified at an Influent Air Flow Rate of 760 m^3(STP)/h, with No Foam Present and No Substrate Limitation.**

Required Oxygen Supplementation Rates kg(O_2)/h To Operate the Aerobic Reactor at 50°C Under Non-Foaming Conditions				
Retention Time (days)	3.0	2.0	1.5	1.0
Winter (15°C)	17	34	51	86
Average (20°C)	10	25	40	70
Summer (25°C)	4	16	28	53
Required Oxygen Supplementation Rates kg(O_2)/h To Operate the Aerobic Reactor at 60°C Under Non-Foaming Conditions				
Retention times (days)	3.0	2.0	1.5	1.0
Winter (15°C)	45	67	89	135
Average (20°C)	38	58	78	118
Summer (25°C)	32	49	67	102

5.4.3 Oxygen Injection Rates to Increase Loading: Foaming Conditions

When foaming takes place on the surface of the sludge in the aerobic reactor, the aeration characteristics of the system improve (see Section 4.3.5). Foaming was only observed in the reactor when the temperature was in excess of 50°C and was unpredictable in its occurrence. It is envisaged that by providing greater control over the reactor temperature with pure oxygen supplementation the foam layer can be sustained and the benefits of foaming can be fully and consistently exploited.

In modelling oxygen supplementation under conditions of foaming, the aeration characteristics of the air aeration system are accepted to be set at the optimum as established in Section 5.3.2 ($\mathbf{Q}\ (AIR)_{in}^{opt}$ = 480 m^3(STP)/h and tabulated in Table 5.3. The steady state heat balance is then calculated in the same manner as described above taking account of the improved *OTE* (23.7%) with foam.

The Biological Heating Rate

$$H_b = H_b^{AIR} + H_b^{O_2} = Y_h.V_p.6.49\times10^{-4}\times Q(AIR)_{in}\times e^{-0.00124Q(AIR)_{in}} + Y_h.OTE^{O_2}.M(O_2)_{in}^{O_2} \quad \text{...MJ/h} \quad (5.79)$$

where:

Y_h	= Specific Heat Yield	12.8	MJ/kg(O_2)
V_p	= Aerobic reactor process volume	184	m^3
$Q(AIR)_{in}$	= Optimum influent air flow rate under foaming conditions	480	m^3(STP)/h
$M(O_2)_{in}^{O2}$	= Mass (pure) oxygen supplementation rate	$M(O_2)_{in}^{O2}$	kg(O_2)/h
OTE^{O2}	= Pure oxygen transfer efficiency	0.80	-

$$H_b = 12.8\times184\times6.49\times10^{-4}\times480\times e^{-0.00124\times480} + 12.8\times0.80\times M(O_2)_{in}^{O_2} = 405 + 10.2M(O_2)_{in}^{O_2} \quad \text{...MJ/h} \quad (5.80)$$

The Mechanical Heat Input Rate

$$H_m = 0.0036 \times f_{mech}.\cos\phi.\sqrt{3}.V.I \quad \text{...MJ/h} \quad (5.81)$$

where:

f_{mech}	= Fraction of electrical energy converted to heat	0.85	
V	= Phase voltage of the pump motor	380	volts
I	= Phase current drawn by the pump motor	19	amps
$\cos\Phi$	= Power factor for the pump motor	0.82	
0.0036	= Conversion factor from W to MJ/h	0.0036	MJ/W.h

$$H_m = 0.0036 \times 0.85 \times 0.82 \times 1.732 \times 380 \times 19 = 31 \quad \text{...MJ/h} \quad (5.82)$$

The Effluent Gas Vapour Heat Loss Rate

$$H_v = L_v^{50^\circ C}.\rho(AIR).Q(AIR)_{in}.(U_{out} - U_{in}) \quad \text{...MJ/h} \quad (5.83)$$

where:

$L_v^{50^\circ C}$	= Latent heat of vaporisation of water at 50°C	2.38	MJ/kg(H_2O)
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	kg(AIR)/m^3
$Q(AIR)_{in}$	= Optimum influent air flow rate under foaming conditions	480	m^3(STP)/h
U_{out}	= Effluent air stream humidity (at 50°C)	0.087	kg(H_2O)/kg(AIR)
U_{in}	= Influent air stream humidity (function of T_{amb})	$f(T_{amb})$	kg(H_2O)/kg(AIR)

$$H_v = 2.38\times1.21\times480(0.087 - (0.00058T_{amb} - 0.0020)) = 123 - 0.80T_{amb} \quad \text{...MJ/h} \quad (5.84)$$

The Effluent Gas Sensible Heat Loss Rate

$$H_g = \rho(AIR).Q(AIR)_{in}.\left(C_p^{20^\circ C}(AIR).(T(AIR)_r - T(AIR)_{in}) + U_{out}.C_p^{50^\circ C}(H_2O)_{vap}.T(AIR)_r\right) \text{...MJ/h} \quad (5.85)$$

where:

$\rho\ (AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	kg(AIR)/m^3
$Q(AIR)_{in}$	= Optimum influent air flow rate under foaming conditions	480	m^3(STP)/h
$C_p^{20^\circ C}$ (AIR)	= Heat capacity of dry air at STP	0.0010	MJ/kg(AIR)
$C_p^{50^\circ C}$ $(H_2O)_{vap}$	=Heat capacity of water vapour at 50°C	0.00187	MJ/kg(H_2O)
U_{out}	= Effluent air stream humidity (at 50°C)	0.087	kg(H_2O)/kg(AIR)
$T(AIR)_{in}$	= Influent air stream temperature	T_{amb}	°C
$T(AIR)_r$	= Effluent air stream temperature	50	°C

$$H_g = 1.21 \times 480(0.0010(50 - T_{amb}) + 0.087 \times 0.00187 \times 50) = 34 - 0.58T_{amb} \text{...MJ/h} \quad (5.86)$$

The Effluent Sludge Sensible Heat Loss Rate

$$H_s = C_p^{20^\circ C}(SL)\left(\left(\frac{Q(SL)_{in}}{24} - \frac{(\rho(AIR).Q(AIR)_{in}.U_{out}}{1000}\right).T(SL)_r - \frac{Q(SL)_{in}.T(SL)_{in}}{24}\right) \text{...MJ/h} \quad (5.87)$$

where:

$C_p^{20^\circ C}(SL)$	= Heat capacity of sludge at 20°C	4.0	MJ/m^3
$Q(SL)_{in}$	= Influent sludge stream flow rate	$Q(SL)_{in}$	m^3/d
$\rho(AIR)$	= Density of dry air at 20°C and 760 mmHg (STP)	1.21	kg(AIR)/m^3
$Q(AIR)_{in}$	= Optimum influent air flow rate under foaming conditions	480	m^3(STP)/h
U_{out}	= Effluent air stream humidity (at 50°C)	0.087	kg(H_2O)/kg(AIR)
$T(SL)_{in}$	= Influent sludge stream temperature	T_{amb}	°C
$T(SL)_r$	= Effluent sludge stream temperature	50	°C

$$H_s = 4.0\left(\left(\frac{Q(SL)_{in}}{24} - \frac{1.21 \times 480 \times 0.087}{1000}\right)50 - \frac{Q(SL)_{in}T_{amb}}{24}\right) = 8.33Q(SL)_{in} - 10 - 0.167Q(SL)_{in}.T_{amb} \text{...MJ/h} \quad (5.88)$$

The Wall Heat Loss Rate

$$H_w = 0.82(T(SL)_r - T(SL)_d) + 0.49(T(SL)_r - T(AIR)_{in}) \text{...MJ/h} \quad (5.89)$$

where:

$T(SL)_r$	= Aerobic reactor sludge temperature	50	°C
$T(AIR)_{in}$	= Ambient air temperature	T_{amb}	°C
$T(SL)_d$	= Anaerobic digeser sludge temperature	$(50+T_{amb})/2$	°C

$$H_w = 0.82(50 - (50 + T_{amb})/2) + 0.49(50 - T_{amb}) = 45 - 0.9T_{amb} \text{...MJ/h} \quad (5.90)$$

Combining each of the above heat terms in the steady state heat balance:

$$H_b + H_m = H_v + H_g + H_s + H_w \quad \text{...MJ/h} \quad (3.8)$$

yields:

$$405 + 10.2M(O_2)_{in}^{O2} + 31 = 123 - 1.80T_{amb} + 34 - 0.58T_{amb} + 8.33Q(SL)_{in} - 10 - 0.167Q(SL)_{in}.T_{amb} + 45 - 0.9T_{amb} \quad \text{...MJ/h} \quad (5.91)$$

Rearranging in terms of $M(O_2)_{in}^{O2}$ yields:

$$M(O_2)_{in}^{O_2} = 0.81Q(SL)_{in} - 0.22T_{amb} - 0.0163Q(SL)_{in}.T_{amb} - 23.9 \quad \text{...MJ/h} \quad (5.92)$$

Which in terms of retention time R_h (note $Q(SL)_{in} = V_p / R_h = 184 / R_h$) *is* given by:

$$M(O_2)_{in}^{O2} = \frac{184}{R_h} \times (0.81 - 0.0163T_{amb}) - 0.22T_{amb} - 23.9$$

Equation 5.93 **The Rate of Pure Oxygen Supplementation Required ($kg(O_2)/h$) to Achieve a Reactor Temperature of 50°C as a function of the Retention Time and Ambient Temperature under Non-Foaming Conditions.**

The variation in the required rate of oxygen supplementation, to allow the aerobic reactor to reach an operating temperature of 50°C, for specific retention times between 3 and 1 days, at different ambient temperatures is illustrated in Figure 5.11 below. Note that this graph is valid for conditions when the reactor **is foaming** and the air flow rate is set at its optimum rate for foaming conditions (480 m^3(STP)/h). For comparison purposes, the oxygen supplementation rates calculated for non-foaming conditions (see Section 5.4.2) are indicated on the graph by the dashed lines (extracted from Figure 5.10).

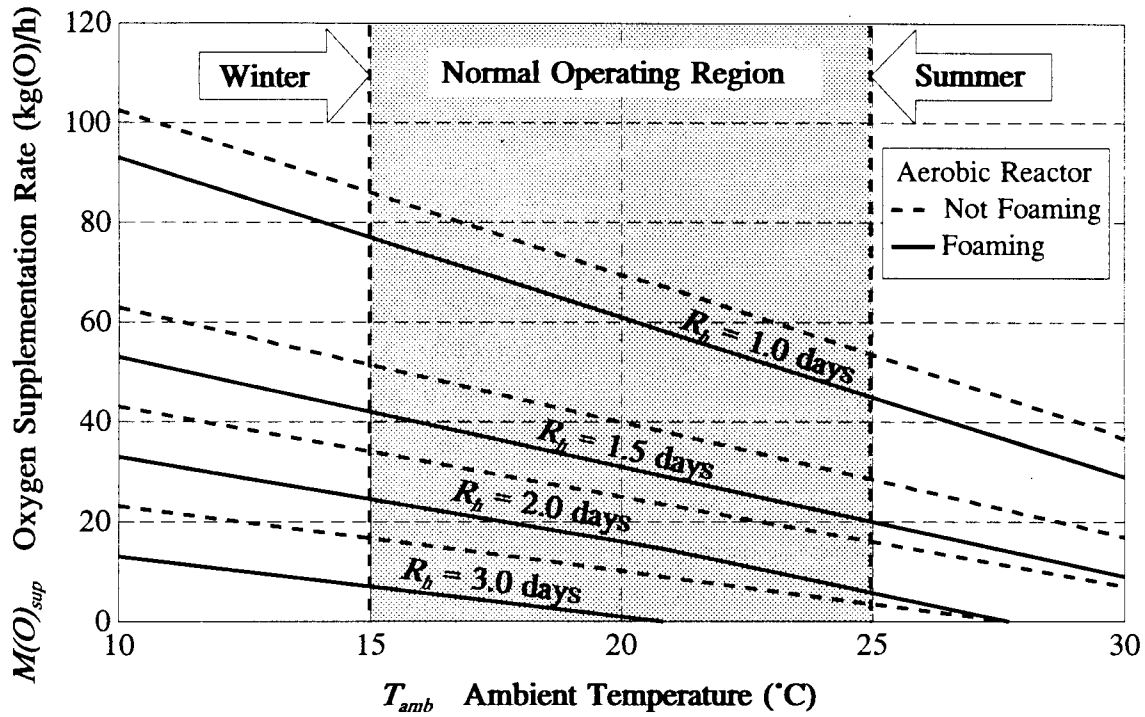

Figure 5.11 The Variation in the Required Rate of Oxygen Supplementation which will allow a Reactor Temperature of 50°C, with Ambient Temperature for 1.0, 1.5, 2.0, and 3.0d Sludge Retention Times: The Air Flow Rate is set at (480 m^3(STP)/h and the Reactor is Foaming.

For design purposes, accepting an average ambient temperature of 25°C in summer and 15°C in winter, Table 5.5 below gives the required oxygen supplementaiton rates to maintain the aerobic reactor at 50°C for different sludge retention times. The data is valid for an influent air flow rate of 480 m^3(STP)/h, the reactor is foaming, and there is no substrate limitation.

In making comparison with oxygen supplementation rates calculated for non-foaming conditions (Table 5.4) it can be seen that the required oxygen supplementation rates are reduced by approximately 8kg(O_2)/h as a result of the foaming; under foaming conditions there is a slight increase in the predicted oxygen utilisation rate for air (0.15 → 0.17 kg(O_2)/m^3.h) which increases the biological heating rate, and the lower air flow rate (480 m^3(STP)/h) means that the vent gas heat losses are reduced. Consequently lower oxygen supplementation rates are required to operate the reactor at a specific temperature (in this case 50°C) when the reactor is foaming.

Table 5.5 The Required Oxygen Supplementation Rates to Maintain the Aerobic Reactor Temperature at 50°C and 60°C for the Sludge Retention Times Specified at an Influent Air Flow Rate of 480 m^3(STP)/h, with Foam Present and No Substrate Limitation.

Required Oxygen Supplementation Rates kg(0_2)/h To Operate the Aerobic Reactor at 50°C Under Foaming Conditions				
Retention Time (days)	3.0	2.0	1.5	1.0
Winter (15°C)	8	26	43	78
Average (20°C)	2	17	32	62
Summer (25°C)	0	8	20	46
Required Oxygen Supplementation Rates kg(0_2)/h To Operate the Aerobic Reactor at 60°C Under Foaming Conditions				
Retention times (days)	3.0	2.0	1.5	1.0
Winter (15°C)	27	50	73	118
Average (20°C)	21	41	61	101
Summer (25°C)	15	33	50	85

The pure oxygen supplementation rates required to operate the reactor at 60°C are also presented in Table 5.5 above. The rates were calculated in the same manner as those for operation at 50°C described above. In making comparison with oxygen supplementation rates calculated for non-foaming conditions for aerobic reactor operation at 60°C (see Table 5.4) it can be seen that the required oxygen supplementation rates are reduced by approximately 17kg(O_2)/h as a result of foaming.

CLOSURE

The models presented in Section 5.4 (Pure Oxygen Supplementation) were developed solely to make an estimate of the rates of pure oxygen injection (pure oxygen supply rates) during phase II of the investigation. This enabled the pure oxygen injection equipment to be properly sized. A more detailed approach to modelling the process (where oxygenation is with air + pure oxygen) is contained in Chapter 8. Use is made of the operational data obtained during phase II and allowance is made for the change in gas stream flow rate and mechanical heat input.

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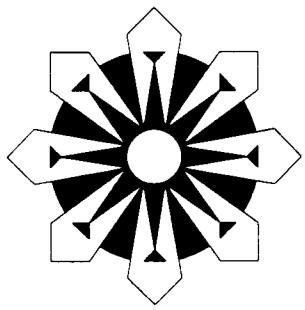

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS: PHASE I

6.1 Overall Assessment of Phase I

Phase I of the study, to investigate the dual digestion process using air to stabilise primary sewage sludge, was successful in that virtually all the initial objectives laid out for the project were achieved. The plant operated continuously for a 312 day evaluation period and no monitoring problems were encountered. Consequently sufficient data was collected to fully evaluate the system. Because of the fundamental importance of biological heat generation to the performance of the aerobic reactor, special emphasis was placed on examining the aerobic reactor. A steady state mass (liquid and gaseous) and heat balance over the aerobic reactor was formulated. The consistency of the results obtained for the specific heat yield (Y_h) and respiration quotient (Y_{CO2}) for 8 different steady state periods under different operating conditions indicated that the mass and heat balance equations were sufficiently general and sensitive to the changing operating conditions. Aerobic reactor process stoichiometry was well defined and good agreement was obtained with previous workers. As a result, with Y_h and Y_{CO2} known, the mass and heat balance equations can be reliably used for design and evaluation of aerobic reactors aerated with air and pure oxygen enriched air. The central and most important information required for design is a knowledge of the oxygen transfer characteristics (such as oxygen transfer rate and efficiency at different air/oxygen supply rates) of the aeration device(s).

6.2 Conclusions Drawn from Phase I

In order to present the conclusions from phase I of the investigation in logical format, firstly those with regard to the aerobic reactor are presented. These are followed by the conclusions drawn with regard to the objectives of the dual digestion system evaluation. The Chapter closes with an assessment of the viability of the air dual digestion system for sewage sludge treatment.

In the synopsis, a summary is given of; (i) the objectives of this investigation; (ii) the dual digestion system layout, operation and monitoring; (iii) the results; (iv) design considerations; (v) evaluation of the air dual digestion system; and (vi) the recommendations for further research. These aspects are not repeated in this Chapter; rather in this Chapter detail conclusions from the investigation itself are presented, under various headings.

6.2.1 Characteristics of the Aerobic Reactor Aeration System

It was accepted for the purposes of this investigation that the oxygen utilisation rate *OUR* was the most appropriate parameter for monitoring the rate of biological heat generation H_b in the thermophilic aerobic reactor in the dual digestion system. To measure the *OUR*, the characteristics of the aeration system needed to be well defined. Measurement of both the carbon dioxide and oxygen volumetric fractions in the effluent gas stream and assuming the nitrogen gas mass flow rate remained constant through the reactor, allowed accurate estimation of the *OUR* and the respiration quotient Y_{CO2}. The effect of foaming on the aeration characteristics was also investigated.

Conclusions:

1.1 The aerobic reactor operated under oxygen limiting conditions during each steady state period (confirmed by the absence of dissolved oxygen in the reactor sludge). Under such conditions, the biological oxygen utilisation rate ***OUR*** was limited by the maximum oxygen transfer rate $\boldsymbol{OTR}_{max}$ which could be effected by the aeration device.

1.2 The maximum oxygen transfer rate which the aeration device could deliver under standard conditions using tap water in the non-steady state aeration test ($\boldsymbol{OTR}^{std}_{max}$) was measured. The value obtained for ($\boldsymbol{OTR}^{std}_{max}$) from the test was 0.167 $kg(O_2)/m^3.h$. The influent dry air flow rate was 760 m^3(STP)/h (the maximum possible delivery rate from one compressor), and the oxygen transfer efficiency ***OTE*** was 14.5%.

1.3 Under non-foaming conditions, with an influent dry air flow rate of 760 m^3 (STP)/h the measured oxygen utilisation rate ***OUR*** remained relatively constant at approximately 0.15 $kg(O_2)/m^3.h$. The average oxygen transfer efficiency ***OTE***, at this air flow rate, was measured to be 13.1%.

1.4 The maximum oxygen transfer rate OTR_{max} which the aeration device could deliver at an flow rate of 760 m^3(STP)/h, under non-foaming conditions at thermophilic temperatures, was approximately 10% lower than the maximum oxygen transfer rate OTR^{std}_{max} measured under standard conditions in the non-steady state aeration test. This difference is in agreement with that claimed by Fuggle and Spensely (1985). The operating and standard test *OTE's* at 760 m^3(STP)/h were 14.5% and 13.1% respectively.

1.5 The presence of a significant foam layer on the surface of the sludge in the reactor improved the oxygen transfer efficiency *(OTE)* from around **13%** at 760 m^3(STP)/h (*OSR* = 1.148 kg(O_2)/m^3.h) to **24%** at 420 m^3(STP)/h (*OSR* = 0.675 kg(O_2)/m^3.h) and **32%** at 220 m^3(STP)/h (*OSR* = 0.345 kg(O_2)/m^3.h).

1.6 The aeration characteristics of the mixing/aeration device were established for both foaming and non-foaming conditions under different operating conditions, viz:

Foaming: $OTR = OSR \times 43.0 \times e^{-0.822 OSR}$ in the range $0.33 < OSR < 0.85$ kg(O_2)/m^3.h

Non-Foaming: $OTR = OSR \times 18.1 \times e^{-0.322 OSR}$ in the range $0.75 < OSR < 2.00$ kg(O_2)/m^3.h

Within the specified ranges the *OTR* generally increases with increases in *OSR*, but increasingly less so due to the decrease in *OTE* as *OSR* increases. The aeration characteristics (ie. the *OTR* - *OSR* relationship) is unique for each aeration device but needs to be known to establish reliable aerobic reactor heat balance relationships.

1.7 Control of the foam layer was effected by reducing the influent air flow rate. Reducing the flow rate lowered the oxygen supply rate *(OSR)* significantly. However, a concomitant decrease in oxygen transfer rate *(OTR)* and hence, oxygen utilisation rate *(OUR)*, was not observed due to the doubling in oxygen transfer efficiency *(OTE)* under foaming conditions.

1.8 The maximum oxygen transfer rate OTR^{std}_{max} calculated from the non-steady state standard aeration test, cannot be used for predicting the maximum oxygen transfer rate OTR_{max} for foaming conditions in the reactor. Under foaming conditions, the contact time between the liquid and gas phases, as well as the interfacial area between the two phases increases markedly. As both of these factors are incorporated into the liquid-gas oxygen mass transfer coefficient K_{La}, an increase

in the K_{La} value occurs under foaming conditions; this increased K_{La} giving rise to an increase in the OTR_{max}.

1.9 From a heat balance point of view, it was not practical to aerate the reactor with two liquid ring compressors (combined air flow rate of 1200 m^3(STP)/h). The ***OTR*** and hence ***OUR*** did not increase in proportion to the increase in the ***OSR***. The former increased from 0.15 to 0.18 kg(O_2)/m^3.h in response to an increase in the latter from 1.16 to 1.82 kg(O_2)/m^3.h. The additional biological heat generated from the increased ***OTR*** could not compensate for the increased vent gas sensible and water vapour heat losses from the increased ***OSR*** and consequently lower reactor temperatures were observed.

1.10 The respiration quotient Y_{CO2} (defined as the number of moles of carbon dioxide generated per mole of oxygen utilised) was measured to be 0.70 and not 1.0 as is often assumed. The measured value of 0.70 is in good agreement with that obtained by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor (0.68).

6.2.2 The Biological Heating Rate and the Steady State Heat Balance

The biological heating rate H_b is of fundamental importance to the performance of the aerobic reactor. Calculation of H_b by difference from the steady state heat balance allows the specific heat yield Y_h to be determined. Once Y_h is known the heat balance can be used to predict the minimum required retention time to reach thermophilic temperature under a variety of different operating conditions.

Conclusions:

2.1 The biological heating rate H_b was directly proportional to the oxygen transfer rate ***OTR*** which, under oxygen limiting conditions, fixes the biological oxygen utilisation rate ***OUR***. The constant of proportionality is the specific heat yield Y_h.

2.2 The specific heat yield Y_h (defined as the quantity of biological heat produced per mass of oxygen utilised) was measured to be 12.8 MJ/kg(O_2). This shows good agreement with the value of 13.0 obtained by Messenger *et al* (1992). The spread in Y_h values between each steady state period was small (12.3 to 13.2 MJ/kg(O_2)) in spite of the widely differing operating conditions during the 8 steady state periods.

2.3 The assumption made in calculating the steady state heat balances that the effluent (vent) gas stream from the aerobic reactor was saturated with water vapour was found to be correct. In the heat balance, it was assumed that the effluent gas from the aerobic reactor was saturated with water vapour. The vapour heat loss rate H_v is one of the major heat loss terms in the heat balance, for the aerobic reactor oxygenated with air, with the result that the calculated Y_h value is very sensitive to the effluent gas stream humidity. The good agreement the calculated specific heat yield Y_h with that of Messenger *et al* (1992) confirms the accuracy of the heat balance and therefore indirectly verifies the assumption that the effluent gas stream is saturated with water vapour. The production of large quantities of condensate observed in the reactor gas headspace provided practical proof of this.

2.4 Significantly higher temperatures (+5°C) were achieved in the aerobic reactor during periods of heavy foaming (foam >3m deep). Foam spillages could only be controlled by reducing the ***OSR***. At reduced ***OSR's*** (420 and 220 m^3(STP)/h) with foaming, similar oxygen transfer and utilisation rates ***(OTR*** and ***OUR)*** were achieved, compared with periods of non-foaming, inspite of the reduction in ***OSR*** to control foam spillage. Accordingly the biological heating rate H_h remained approximately constant but reducing the influent air flow rate reduced the effluent gas sensible (H_g) and water vapour (H_v) heat losses significantly. With the reduced heat losses, under foaming conditions the reactor temperature increased about 5°C to 55°C.

6.2.3 Aerobic Reactor Retention Time

Owing to the high effluent gas heat losses with air aeration, the retention time of the aerobic reactor needed to be considerably longer (4-6d) than that for the pure oxygen aerobic reactor (1.25 to 1.5d)

Conclusions:

3.1 In the winter when the feed sludge temperature was around 15°C, the aerobic reactor retention needed to be around 6 days to maintain a reactor temperature of 50°C.

3.2 In the summer, at a feed sludge temperature of 25°C, the retention time could be reduced to 3.8 days to maintain a reactor temperature of around 50°C

6.2.4 Volatile Solids Destruction in the Aerobic Reactor

A significant degree of volatile solids destruction takes place in the aerobic reactor as a consequence of the relatively long retention times employed. As a result it was possible to quantify this reduction in terms of the biological heating rate and the oxygen utilisation rate. For the pure oxygen aerobic reactor at Milnerton operated at very short retention times, this was not possible because the VS removal was very low (<1.5%) (Messenger *et al*, 1992).

Conclusions:

4.1 At the long retention times of the aerobic reactor, a significant degree of volatile solids destruction took place (**25%** ie 9.3 kg(VS)/m^3 from 37kg(VS)/m^3 average influent concentration at 41m^3/d, 4.4d retention time).

4.2 The quantity of biological heat produced per mass of volatile solids destroyed (MJ/kg(VS)) was 22. This shows good agreement with the value of 21 obtained by Andrews and Kambhu (1971) which has become a recognised standard value for the design of the ATAD process.

4.3 The quantity of oxygen consumed per mass of volatile solids destroyed was calculated at 1.70kg(O_2)kgVS. This figure is higher than that normally associated with sewage sludge (1.42 mgCOD/mgVS) but is in agreement with the COD/VS ratio of the influent feed sludge.

4.4 The average mass oxygen utilisation rate $M(O_2)_{ut}$ calculated over the full investigation period was 27.6 kg(O_2)/h. This figure is 12.4% lower than the calculated average rate of COD destruction (31.5 kg(COD)/h) which took place in the aerobic reactor. This difference makes the COD balance 87.6% and may be attributable to the variability associated with the COD test. If this is so, then it can be assumed that the COD balance over the reactor was maintained (unlike the Milnerton pure oxygen reactor where the COD balance was 234% ie the oxygen mass consumed was 2.34 times greater than the COD mass removed).

4.5 From the VS destroyed (25% of 37kgVS/m^3 at 41m^3/d), 206.4kgC/d were mineralised based on a $C_5H_7O_2N$ formula for sludge. From the average oxygen utilisation rate 27.6kg(O_2)/h and respiration quotient Y_{CO2} of 0.70 mole CO_2 produced per mole O_2 utilised, 174.3kgC/d carbon dioxide escaped in the vent gas. From the increases in alkalinity and ammonia concentration, approximately 8.9kgC/d of carbon dioxide remained dissolved in the liquid phase and escaped

with the effluent sludge flow. This meant that 183.2 kgC/d of carbon dioxide can be accounted for which represents 89% of the carbon mineralised. Clearly, at Athlone a reasonably good carbon balance was obtained, unlike at Milnerton where the carbon mineralised on the same $C_5H_7O_2N$ basis (7.5 kgC/d) was only 7.5% of the carbon in the vent gas (100 kgC/d), giving a carbon balance of 1300%!. Full details of the carbon balance are provided in Appendix 6.

4.6 The above results permit an integrated mathematical model to be developed linking biological heating rate, oxygen utilisation rate, %VS removal to retention time and reactor temperature. The success of the model is discussed in Section 6.2.8 below). From the %VS removal, the increase in ammonia concentration could be estimated (see Section 6.2.5 below).

6.2.5 Conditioning Effects of Aerobic Pre-Treatment

Objective 1 Conditioning: To assess the conditioning effects of aerobic pre-treatment on; (a) reducing the minimum required time for subsequent anaerobic digestion which would increase digester capacity, and (b) providing greater pH stability to the digestion stage.

Conclusions:

5.1 As a result of aerobic treatment the average ammonium and bicarbonate alkalinity concentrations increased from 113 mgN/l and 40 mg$CaCO_3$/l in the feed to 365 mgN/l and 820 mg$CaCO_3$/l in the reactor respectively. The pH increased from an average 5.4 to 7.4.

5.2 Because the anaerobic digester had to be operated at its full operating capacity (1800 m^3) to ensure good mixing, the effect of aerobic reactor heat conditioning on reducing the minimum required retention time for anaerobic digestion could not be tested during phase I; this was subsequently done at laboratory scale and during phase II where pure oxygen supplementation made higher loading rates possible.

5.3 The laboratory scale study demonstrated the viability of operating the anaerobic digester at an 8 day retention time. Whilst the mesophilic (37°C) digester (at 8d retention time) fed with primary sludge failed soon after start up, an identical digester fed with aerobically treated sludge remained operating stably. It was concluded that the stability of the anaerobic digester biological processes is a criterion of much less significance in establishing the anaerobic digester retention

time than the stability of the final sludge product (in terms of %VS removal and/or specific oxygen utilisation rate ***SOUR*** < 1.0 $g(O_2)/kgTSS.h$).

5.4 Both the ammonium ion concentration and the bicarbonate alkalinity produced in the aerobic reactor could be accurately estimated from the quantity of volatile solids destroyed. Theoretical predictions could also be made regarding the increase in alkalinity for different retention times.

5.5 No nitrification took place in the aerobic reactor; had this taken place the alkalinity would have been reduced to virtually zero because the alkalinity generation arises principally from the ammonification of proteins ($NH_3 \rightarrow NH_4$).

5.6 Approximately 5% of the carbon dioxide generated through the mineralisation of organics remains in solution and reacts with the ammonium ion to form ammonium bicarbonate which increases the alkalinity.

5.7 Because of the long retention times required for operation with air the degree of volatile solids destruction which takes place is significant. It may be more appropriate to regard the sludge as part-stabilised rather than conditioned.

5.8 Although not examined on the anaerobic digester, the high VS removal (part-stabilisation) in the aerobic reactor would in all likelihood reduce digester gas production due to the reduced VS solids load per unit flow.

6.2.6 Requirements for Disinfection and Heating the Anaerobic Digester

Objective 2 Disinfection: To demonstrate that sufficiently high temperatures can be achieved in the aerobic reactor, with a typical sewage sludge, using a simple aeration system such that (a) a satisfactory degree of disinfection is achieved and that (b) the subsequent heat requirements for optimum mesophilic anaerobic digestion are met.

Conclusions:

6.1 Sufficient autoheating potential existed for thermophilic temperatures (>50°C) to be maintained throughout the year. The estimated minimum retention time for 50°C operation in summer (25°C) is 3.6 days and in winter (15°C) is 5.8 days. However, the the actual retention times required will be entirely dependent on the ***OTR-OTE*** characteristics of the aeration system (under oxygen limiting conditions, which is usually the case).

6.2 There is a minimum feed sludge concentration, which increases with increasing retention time, which must be maintained to ensure that substrate limitation does not take place and that instead the reactor operates under oxygen limiting conditions. Under oxygen limiting conditions the biological heating rate is fixed by the aeration system oxygen transfer rate (***OTR***).

6.3 Disinfection in terms of 100% inactivation of Ascaris ova was achieved in the reactor when the temperature was in excess of 50°C. With no evidence of short circuiting, it can be concluded that the design of the feed pattern and mixing pattern was adequate. (It will however be recommended that future reactors should be designed to operate on a draw and fill basis rather than a displacement flow through batch basis). At temperatures below 50°C, 3% of the *Ascaris* ova were viable in the reactor effluent.

6.4 An average of 3 orders of magnitude reduction in faecal coliforms was observed across the aerobic reactor (from 9.3×10^8 to 7.5×10^5/100ml at 3.7% and 2.8% volatile solids concentration respectively) with a further 1 order of magnitude reduction after anaerobic digestion (to 4.3×10^4/100ml at 1.7% volatile solids concentration). These reductions are better than the $2\log_{10}$ reductions required for Processes that Further Reduce Pathogens (PFPR).

6.2.7 Quality of the Final Sludge

Objective 3: To assess the quality of the final sludge in terms of stability (%VS removal), fermentability (gas production), odour and dewaterability (specific resistance to filtration); and to compare these with conventional anaerobic treatment.

Conclusions:

7.1 The final sludge from the digester was stable and did not undergo further fermentation. The volatile solids content of the final sludge was 70% (dry mass basis). The average percentage VS, TS and COD removals in the aerobic reactor were 25, 23 and 33% respectively and in the anaerobic digester were 40, 32 and 40% respectively. This was after an overall average of 4.6 days in the aerobic reactor and 42 days in the anaerobic digester.

7.2 The average percentage VS, TS and COD removals in the system overall were 56, 48 and 59% respectively and are comparable to conventional anaerobic digestion.

7.3 A stable sludge is regarded as one on which a %VS removal in excess of 38% has been achieved, and has a residual specific oxygen utilisation rate ***SOUR*** of <1g(O)/kg(TSS).h. With respect to the %VS removal the sludge would be regarded as stable, but because the ***SOUR*** was not measured during phase I (it was during phase II), its compliance with the ***SOUR*** criterion is not known.

7.4 In the 8d retention time laboratory scale digester fed with aerobic reactor sludge, the VS removal was 46% and the volatile solids content of the effluent sludge was 74%. This VS removal is actually higher than the 40% VS removal achieved in the full scale digester which operated at 42 days. This indicates that retention times in excess of 8 days are probably longer than required for sludge stabilisation, but residual ***SOUR*** would need to be tested to check this! The superior performance of the laboratory scale digester, in terms of VS removal, is likely to be due to the higher operating temperature (37°C as opposed to 31°C) and more efficient mixing. If short retention times are to be employed at full scale then the operating temperature should be kept as close as possible to the optimum of 37°C and mixing should be efficient. If this cannot be achieved then it would be advisable to operate at longer retention times to safeguard against digester upset.

7.5 The final sludge from the digester had an earthy odour which was identical to that from conventional anaerobic digestion. The sludge from the aerobic reactor (as far as could be subjectively established from operating staff) also did not have an offensive odour.

7.6 The dewaterability of the final sludge (SRF = 368 x 10^{12}m/kg) was not significantly different to that of the conventional anaerobically digested sludge at Athlone (218 x10^{12}m/kg) or the pure oxygen dual digestion sludge at Milnerton (507 x 10^{12}m/kg) (see also Messenger *et al*, 1992).

6.2.8 Defining the Aeration and Heating Requirements to Achieve Autoheating

Objective 4: To define the aeration and heating requirements to achieve autoheating with air oxygenation and to establish the minimum practical aerobic and anaerobic retention times for the process. To develop a mathematical model to enable prediction of reactor temperatures and to recommend criteria for the design and operation of future plants.

Conclusions:

8.1 The investigation was successful in defining both the aeration and heating requirements to achieve autoheating with air oxygenation.

8.2 A mathematical model based on the calibrated and verified steady state heat balance was developed which was capable of predicting reactor temperatures under a variety of conditions or alternatively the minimum retention time to operate at a specified thermophilic temperature for different sludge feed and ambient temperatures.

8.3 The model was sufficiently accurate to place confidence in the recommended criteria for the future design of aerobic reactors either oxygenated with air alone or a combination of air and pure oxygen (pure oxygen supplementation).

6.2.9 Operation of the Dual Digester from a Practical Viewpoint

Objective 5: To estimate the practical operational problems of a full scale system over an extended period and to make an assessment of its reliability.

9.1 The process proved to be simple to operate. The number of staff required to operate the dual digestion plant would be no different from that required for conventional anaerobic digestion.

9.2 The aerobic reactor was easily started. From a starting temperature of 18°C the reactor reached thermophilic temperatures within 10 days. Typically the rate of temperature increase was 3°C per day.

9.3 It is important for the operating staff to ensure that the concentration of feed sludge to the reactor is adequate to prevent substrate limitation taking place.

9.4 The operating staff need to constantly monitor the foam layer in order to prevent spillage. Foam layers in excess of 3m can be obtained. With regard to foam control, positive foam management proved successful, by reducing the influent air flow rate to balance the foam production with its collapse.

9.5 Conventional antifoaming practices, ie medium pressure (1atm) water sprays, silicone antifoaming agents (30ppm), and removal by submersible pump suspended in the foam layer, proved unsuccessful for controlling the foam. Foam overflow pipes (150mmϕ) from the top of the aerobic reactor to the transfer boxes of the

anaerobic digester (see Figure 2.2) were not successful in that these functioned satisfactorily only when foam spillage was minor, ie once foam was being controlled via the air flow rate.

9.6 Foam spillages were difficult to clean up. Water jetting and draining proved unsuccessful. Foam could only be removed once it had collapsed due to desiccation and return to a more fluid form, which took on average 3 to 6 hours.

9.7 Whilst routine operation of the aerobic reactor under non-foaming conditions is relatively reliable, the occurrence and subsequent collapse of the foam layer could not be accurately predicted and therefore could not be totally relied upon to provide consistently enhanced performance of the aerobic reactor. However, it appeared that significant foaming only occurred when the reactor temperature exceeded 50°C. To date, little information has been reported on the causes of foaming and how best to promote and exploit this phenomena in aerobic reactors of dual digestion and ATAD systems. In the light of this, and the fact that its occurrence and collapse during this evaluation was unpredictable, an ancillary investigation into the causes of foam formation was undertaken by Samson (1995) at pilot plant scale.

9.8 No major mechanical problems were encountered during the evaluation. The two 7.5kW mixing pumps (used alternately each week) on the sludge recirculation line to the aerobic reactor drew more than their rate of current due to the low head loss through the recirculation loop. The flow rate and hence power required were high, 10kW instead of 7.5kW. Nevertheless, both pumps managed to remain in operation during the evaluation period but the motors burnt out soon after the phase I evaluation period ended.

9.9 On one occasion during the evaluation period, the biological activity in the aerobic reactor was inhibited by a high dosage of cadmium which entered the plant with the raw sewage. The level of cadmium was such, however, that most biological systems would have been adversely affected, verified by the fact that the activated sludge plant lost nitrification at this time.

6.2.10 Capital, Operational, and Maintenance Costs

Objective 6: To estimate the capital, operation and maintenance costs for the system and to make comparisons with conventional anaerobic digestion and the dual digestion system using oxygen.

The Capital Costs: Dual Digestion Using Air

The upgrading of a conventional anaerobic digestion plant to dual digestion is a relatively simple task involving the addition of an appropriately sized aerobic reactor with the necessary set of mechanical equipment to drive the process. The principal capital cost involved in the upgrade therefore will be based primarily on the required size of the aerobic reactor. This can be determined from an application of the steady state heat balance (see Chapter 5) and will be dependent upon:

- The required reactor operating temperature.
- The range of sludge feed and ambient temperatures.
- The volume of sludge to be treated.
- Whether pure oxygen, air, or oxygen enriched air is used for aeration.
- The process volume of the existing anaerobic digester.
- The *OTR-OTE* characteristics of the oxygenation device.

From the results of this particular investigation, to operate and maintain a reactor temperature of 50°C during the winter period with the feed sludge (ambient) temperature at 15°C, the aerobic reactor retention time is required to be at or around 6 days (see Section 6.2.3) and the anaerobic digester retention time at around 15 days to ensure a sufficiently stable sludge product.

Building the aerobic reactor inside the digester with the liquid sludge level equal in both, as in the case with the Athlone system, then the diameter of the reactor is given by $d_{reac} = d_{dig}/\sqrt{(R_{dig}/R_{reac}+1)}$ where R_{reac} and R_{dig} are the retention times of the reactor and digester respectively.

Conclusions:

10.1 For air oxygenation, fixing the retention time (or aspect) ratio R_{dig}/R_{reac} at 3 results in the diameter of the reactor being half the diameter of the digester with the system retention time at 4 times the aerobic reactor retention time.

10.2 With an overall system volume of 2000m^3 (like the Athlone dual digester) the average treatment capacity (20°C, R_{reac} = 5 days) is 2000/(4 x 5) =100m^3/d.

For a pure oxygen system, the aerobic reactor retention times are around 1.5 to 2 days. With anaerobic retention times of 15 to 20 days, the retention time (or aspect) ratio (R_{dig}/R_{reac}) is 10:1 and the overall system retention time is 11 times that of the aerobic reactor retention time.

Conclusions:

10.3 For an overall system volume of 2000m^3 the average treatment capacity with pure oxygen (R_{reac} = 1.5days) is 2000/(11 x 1.5) = 121m^3/d, which is only 20% higher than with air. This demonstrates that the design of the retention time ratio must be suited to whether air or oxygen or oxygen enriched air is used for aeration.

10.4 For the Athlone dual digester, the retention time ratio is suited to pure oxygen (10:1) and this is the principal reason why its treatment capacity was only 40m^3/d ie 40% of the optimum. In the conclusions of phase I, it was recommended (see Section 6.3) that evaluation of the dual digestion system be continued with a phase II investigation. The principal objective being to examine the feasibility of supplementing the aeration system with pure oxygen injection to derive greater benefit from its 10:1 pure oxygen based aspect ratio.

10.5 The capital cost would not be significantly different for a 10:1 or 3:1 ratio dual digester but for the 3:1 air oxygenated system, 3 times more aeration and mixing power would need to be installed and supplied because the aerobic reactor is about 3 times larger than for pure oxygen..

Operational and Maintenance Costs: Dual Digestion Using Air

The recurring running cost for operating the dual digestion process using air is based almost entirely on the electrical power requirements for (1) supplying air to the aerobic reactor and (2) providing mixing in both the aerobic reactor and anaerobic digester. The power rating of the mixing pumps and compressors for the plant at Athlone are as follows:

Table 6.1 Power Rating of the Mechanical Equipment: Athlone Dual Digester

Mechanical Equipment	Power Rating
Liquid Ring Compressor for Air Supply	20 kW
Liquid Ring Compressor for Biogas Recirculation (gas mixing of the anaerobic digester)	20 kW
Aerobic Reactor Mixing Pump	7.5 kW

During the evaluation period the system treated an average sludge flow of 40 m^3/d. Accepting a power cost of 10 cents/kWh the power operating costs are estimated to be 47.5 x 24 x 0.10 /40 = R2.85/m^3 (1990).

Conclusion

10.6 The approximate power running cost of the air dual digestion plant at Athlone was R2.85 per m^3 of sludge treated, at an average aerobic reactor temperature of 50°C.

For comparison with the pure oxygen dual digestion plant at Milnerton, Laubscher *et al* (1992) reported a running cost of R5.29 per m^3 of sludge treated (based on an oxygen cost of 30 cents/kg), at an average aerobic reactor temperature of 60°C. The operational cost of a 3:1 air oxygenated dual digester would be about the same as for the Athlone plant because even though the 3:1 plant would treat 2½ times more sludge, 2½ times more aeration and mixing energy would be required.

Conclusion

10.7 The running cost of operating the dual digestion process using air is approximately 54% that of the process using pure oxygen.

It should be noted however that because of the significant cooling effects inherent when using air, retention times are significantly longer (3-6 days) in comparison to the pure oxygen system (1-1.5 days) and consequently larger sized aerobic reactors are required. The saving in operating costs is therefore partly offset against higher capital costs. In addition, it is possible to maintain higher temperatures in the aerobic reactor throughout the year, ie ~60°C when using pure oxygen as opposed to only 50°C when using air (due to the cooling effects of nitrogen and the need to control the foam level to prevent spillage).

From a maintenance point of view, the performance of the mechanical equipment was routinely monitored and all instrumentation was checked, and re-calibrated if necessary, on a regular basis. The plant did not suffer from any major mechanical problems during the 312 day (10 months) evaluation period.

Conclusion

10.8 The costs associated with personnel and maintenance are not considered to be significantly different from that of a conventional anaerobic digestion plant.

6.2.11 The Viability of the Dual Digestion Process using Air

The fundamental factors in deciding whether or not a particular treatment system is viable for wide application are (1) is it capable of producing a final sludge product that conforms to the criteria laid down for disposal? (2) the ease of operation and its reliability, (3) the cost factor (how competitive is it, with regard to capital, operational and maintenance costs, in comparison with other available systems?), and (4) can the system be readily integrated into the existing treatment plant infrastructure?

Conclusions

11.1 The reactor temperature is probably not consistently high enough throughout the year, to enable the sludge produced to be classified as a type D sludge in terms of the guidelines for the use of sewage sludge (DNH&PD, 1991).

11.2 The system proved easy to operate and reliable. The foaming phenomenon could not be readily exploited as its occurrence could not be predicted except that foaming only occurred with the reactor temperature above 50°C.

11.3 Operating costs for the system are (1) reasonable, (2) about half of that for pure oxygen per m^3 sludge treated, and (3) compare very competitively with other pasteurisation processes.

11.4 For upgrading overloaded digesters, or to provide a pre-pasteurisation stage, a relatively large size reactor would be required to accommodate the retention times needed to allow operation in the thermophilic temperature range with air oxygenation. The effect of ambient temperature on the required retention time to achieve a specific temperature is significant (see Figure 5.2) and therefore the reactor needs to be sized for operation during winter temperatures.

11.5 The retention time in the anaerobic digester needs to be sufficiently short enough (10-15 days) to ensure that the sensible heat contained in the hot sludge from the aerobic reactor would be sufficient to raise the digester temperature to optimum mesophilic temperatures (35-38°C). However, it must not so short as to cause digester overheating if no heat exchanger is available between the reactor effluent and digester influent sludge streams.

11.6 The feed sludge must be have an adequate solids concentration to ensure that substrate limitation does not take place in the aerobic reactor. Sludge pre-thickening greatly benefits the system in terms of kg sludge treated as oxygen requirements are based on sludge volume treated.

11.7 The aeration device employed must be capable of a high oxygen transfer efficiency to ensure that biological heat is not generated at the expense of excessive vent gas heat losses. The ***OTR-OTE*** characteristics of the aeration device must be defined so that heat balance calculations can be made to estimate the oxygen/air requirements.

11.8 Because of the relatively long retention times required to operate the dual digestion system using air in comparison with the system using pure oxygen, a significant degree of VS destruction takes place (25%), with the result that the sludge should be considered partially stabilised rather than pre-treated before anaerobic digestion.

11.9 It is not considered a viable proposition to construct a completely new (air) dual digestion plant (i.e both aerobic reactor and anaerobic digester). By extending the retention time in the aerobic reactor to, for example, 8-10 days the sludge could be stabilised in the aerobic reactor and the anaerobic digester would not be required; ie the reactor would operate as an autothermal thermophilic aerobic digester (ATAD); a system which has gained in popularity in Europe in recent years.

11.10 Using air alone for oxygenation, the Athlone dual digester is capable of handling 30-50 m^3/d of sludge depending on ambient temperature. Treatment capacity can however be increased by increasing the heat input to the aerobic reactor. This can be achieved by (1) pre-heating the feed sludge, and/or (2) supplementing the air oxygenation system with pure oxygen injection. To achieve greater treatment capacity in the Athlone system with its 10:1 aspect ratio, it is recommended that:

- The feasibility of pure oxygen supplementation be investigated. This was undertaken in a phase II investigation. The aims and objectives of this investigation are listed in Section 6.3 below.

- The viability of pre-heating the reactor feed sludge through the heat exchanger system, used for heating the conventional anaerobic digesters, be investigated. This was not undertaken and therefore not documented in this report.

6.3 Recommendations for Future Operation of the Athlone Dual Digestion Plant

6.3.1 Pure Oxygen Supplementation

From the results and conclusions drawn from phase I, it was recommended that further investigative work be carried out on the Athlone Dual Digestion Plant, to evaluate the feasibility of supplementing the aeration of the aerobic reactor with pure oxygen. It was envisaged that with pure oxygen supplementation a substantial proportion of the 200m^3/d primary sludge produced at Athlone could be pasteurised and stabilised by the process without adversely affecting biogas production, which is required as fuel for the gas engine. Consequently, the investigation was extended (called phase II) in order to pursue the above recommendion.

6.3.2 Aims and Objectives of Phase II: Oxygenation with Air + Pure Oxygen

The principal aim of phase II of the investigation was to (a) assess the viability of the system to satisfactorily treat a high proportion (up to 180m^3/d out of 200m^3/d) of the gravity thickened sludge produced at Athlone at a 1 day aerobic reactor retention time and (b) to determine the annual operating costs of this upgraded system. Specific objectives of the investigation were to:

1 **Oxygenation Characteristics:** To determine the pure oxygen supply rate(s) required for a range of different operating conditions, viz:

- Sludge feed volume (90-180m^3/d)
- Aerobic reactor temperature (50-60°C)
- Feed sludge and ambient temperature (15-25°C)
- Retention Time (2-1 day)

and to determine the oxygen transfer efficiencies which can be effected by the pure oxygen injection equipment at the different oxygen supply rates, and the interaction of the air and oxygen oxygenation systems.

2 **Foaming:** To assess the influence which foam formation has on reducing the required pure oxygen supply rate(s). To establish the optimum conditions for foam formation and to determine the requirements for operation with a stable foam layer.

3 **Digester Heating Requirements:** To establish the seasonal heating requirements for optimum operation of the anaerobic digester, i.e. the desired aerobic reactor temperature and retention time which will enable the anaerobic digester to operate at 36-39°C (whilst this was the initial objective it was found that the digester operated quite adequately in the thermophilic temperature range 50°-60°C, and accordingly the heating requirements objectives were ammended).

4 **Costs:** To determine the capital, operation and maintenance costs associated with pure oxygen injection and to make comparison with conventional anaerobic digestion and the dual digestion system using pure oxygen alone.

5 **On Site Oxygen Generation:** To assess the viability of employing a Vacuum Swing Adsorption (VSA) plant for generating oxygen on site. This would include the sizing of an appropriate VSA plant.

6 **External Heating:** To examine the feasibility of pre-heating the feed sludge, and then if possible, determine the reduction in oxygen supply rate which can be effected by pre-heating.

A detailed discussion motivating the phase II investigation and a working programme for it are presented in Appendix 9. The results from the investigation are presented in Chapter 7. Modelling the aerobic reactor performance with air and pure oxygen oxygenation and dual digestion system evaluation are given in Chapters 8 and 9 respectively.

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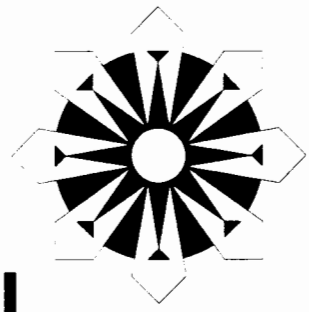

CHAPTER 7

RESULTS AND DISCUSSION: PHASE II

7.1 INTRODUCTION

The general aim of phase II of the research programme was to demonstrate the practicability of the dual digestion process **employing a combination of both air and pure oxygen** to successfully stabilise and disinfect sewage sludge at short anaerobic digester retention times.

It was envisaged, from the conclusions drawn from phase I (see Chapter 6) where operation was with air alone, that the incorporation of pure oxygen supplementation would significantly increase the treatment capacity of the Athlone Dual Digestion facility (i.e. increased sludge throughput and reduced retention times) and provide better process control. Improved process control was envisaged as increased reactor temperature through the use of pure oxygen and improved oxygen transfer rate by controlling the foam layer with the air supply rate. In this way the reactor temperature would no longer be constrained by the capriciousness of the foam layer.

The specific objectives of phase II (see Section 1.7) were planned to cover all the claimed benefits of the dual digestion process and to verify the predictions made for pure oxygen supplementation in Chapter 6. To satisfy these objectives the following aspects of the dual digester performance were evaluated in detail:

- **The utilisation of oxygen within the aerobic reactor**; including measurement of the separate contributions from the air and pure oxygen streams.
- **The aerobic reactor heat balance**; verification of the value determined for the specific heat yield coefficient Y_h during phase I
- **The conditioning effects of aerobic pre-treatment**; As higher aerobic reactor hydraulic loading rates were possible during phase II, the retention time in the anaerobic digester could be reduced to below 10 days.

- **The performance of the anaerobic digester;** In view of the short digester retention time, the stability of the anaerobic process and its sludge product were closely monitored. Due to the high sensible heat content of the sludge transferred from the aerobic reactor, the digester operated at thermophilic temperatures.
- **The dewaterability of the final sludge;** Measured in terms of the specific resistance to filtration test.
- **The bacteriological quality of the final sludge;** Measured in terms of faecal coliform and viable *ascaris* ova concentrations.
- **The stability of the final sludge;** Measured in terms of VS and COD reduction and in terms of the residual Specific Oxygen Utilisation Rate (***SOUR***) over a fixed period of time.

Monitoring of the dual digester during phase II commenced on the 16th July 1994 (designated day 1 of the evaluation period) and lasted for 22 weeks until the 14th December 1994 (day 152). During this period the aerobic reactor sludge recirculation line was out of commission between days 41 and 73 to allow the pipework to be modified[1]. Consequently no pure oxygen injection could take place during this period. The feed sludge flow rate was reduced to a level which allowed the reactor contents to remain active with oxygenation undertaken by the air stream. Outside of this period the plant operated continuously, without any major mechanical problems. Details relating to process reliability are discussed in Section 7.2.2 below.

A summary of the results obtained during phase II are presented and discussed in this Chapter. The Chapter opens with an outline of overall system performance followed by detailed discussion on specific aspects of aerobic reactor and anaerobic digester performance. A complete compilation of the results is given in Appendix 5.

As in phase I, the aerobic reactor of the dual digester was operated for prolonged periods of time under differing steady state conditions to enable reliable data for the determination of the oxygen utilisation rate and consequently the biological heating rate to be obtained to allow solution of the steady state heat balance. The results obtained

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[1] After the initial period of operation (days 1-40) it became evident that an increased pressure and flowrate in the recirculation line was necessary in order to achieve the desired pure oxygen transfer efficiency (>80%). Consequently the recirculation pipeline was modified to achieve this (see Section 7.3.3).

during these steady state periods provided the necessary information to firstly satisfy the above objectives and secondly to expand the calibration of the mathematical model, developed to simulate the performance of the aerobic reactor using air alone (see Chapter 5), and now extended to incorporate pure oxygen supplementation (see Chapter 8).

7.2 SUMMARY OF OVERALL PLANT PERFORMANCE: PHASE II

7.2.1 Introduction

The average performance of the dual digestion plant during phase II, excluding the period during which the recirculation pumps were out of commission (days 41-73) is summarised in Table 7.1 below. The variation in the main operating parameters during this period is shown in Figure 7.1. The shaded areas in Figure 7.1 depict the steady state periods of operation which are discussed in greater detail in Section 7.3 below.

Table 7.1 Summary of Overall Plant Performance During Phase II: Oxygenation of the Aerobic Reactor with both Air and Pure Oxygen.

Sludge Type	Feed Sludge		Aerobic Sludge		Anaerobic Sludge	
Parameter	mean	range	mean	range	mean	range
Temperature °C	20	16-26	57	41-64	43	33-54
Total Solids g/ℓ	47	33-67	39	25-54	27	9-63
Volatile Solids g/ℓ	38	26-55	30	20-44	18	6-34
COD g(O)/ℓ	64	42-93	46	30-69	29	11-61
pH -	5.4	4.8-6.2	7.2	6.7-7.9	7.2	6.7-7.6

Process Stage	Aerobic Reactor		Anaerobic Digester	
Parameter	mean	range	mean	range
Sludge Retention Time d	1.6	2.1-0.95	15.8	9-20
Air Flow Rate m^3/h	333	0-700	-	-
Oxygen Flow Rate kg(O)/h	50	24-96	-	-

%VS Removal	Aerobic	Anaerobic	Overall
	20.2	39.5	51.7

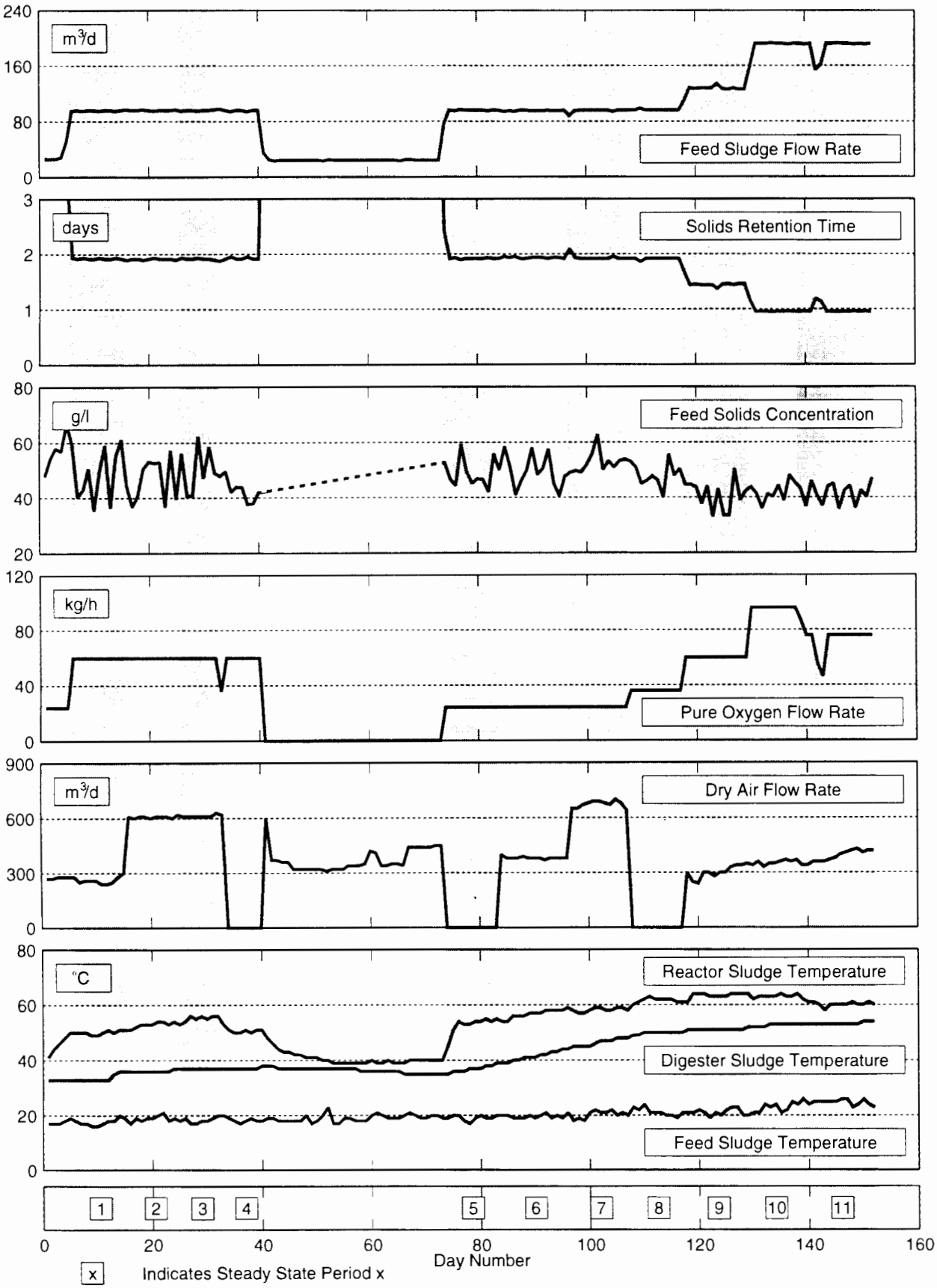

Figure 7.1 **Graphical Representation of the Main Operating Parameters During Phase II**

A summary of the overall plant performance during phase II is discussed below, firstly in terms of the three major system operating parameters (process reliability, temperature and retention time), followed by a brief discussion on the effect of the process on the major chemical characteristics of the sludge.

7.2.2 Process Reliability

As was the case during phase I, the process again proved simple to operate and no major mechanical problems were encountered during the evaluation period. The modifications which were made to the recirculation line (between days 41-73) were carried out in order to effect an improvement in the pure oxygen transfer efficiency. Subsequent to this a variable speed drive, fitted to one of the two recirculation pump motors (see Section 7.3.3), tripped occasionally due to flies blocking the inlet to the cooling fans fitted to the drive. In each instance the pump was restarted after cleaning the inlets. The problem was resolved by fitting a course cloth filter across the inlet to the fan.

Operating conditions in the aerobic stage were generally stable. The only major interruption to the planned operation occurred during steady state period 11, when the pure oxygen storage tank emptied during a weekend and no pure oxygen was available until delivery the following Monday (days 141-142). In order to maintain the reactor temperature the feed sludge flow rate was reduced during this period.

7.2.3 Aerobic Reactor Operating Temperature

The average reactor temperature recorded during phase II was 57°C. The maximum and minimum recorded temperatures were 64°C and 41°C respectively (see Table 7.1). The variation observed in reactor temperature was largely due to the enforced changes made to the air flow rate, pure oxygen flow rate, and the feed sludge flow rate in order to achieve the different steady state conditions (see Figure 7.1). As with the case during phase I (see Section 4.2.3), this variation is not indicative of process instability, but rather a consequence of the changes made to the specific process parameters for research purposes; the effect of the different operating conditions on the reactor temperature is governed by the steady state heat balance, which in turn is influenced by the air, oxygen and sludge influent flow rates. If these three input variables are not changed, the reactor operates stably and achieves a constant temperature[2].

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[2] For the Athlone aerobic reactor, the steady state temperature was a very closely constant value because the reactor was fed on a semi-continuous flow through basis, where the displaced sludge per feed batch represented a small fraction of the reactor volume. As the retention time decreased so the feeding periods became longer and

The seasonal effects due to temperature changes in the feed sludge were not noticeable during phase II as they could be overcome by increasing the pure oxygen flow rate (which increased the biological heating rate); an option not available during phase I where the maximum loading rate was heavily influenced by the ambient temperature due to the oxygen transfer limitations of the air oxygenation system (see Section 5.2.2).

7.2.4 Anaerobic Digester Operating Temperature

At the start of the evaluation period (day 1) the sludge in the anaerobic digester was at a temperature of 33°C. With the digester receiving hot sludge from the aerobic reactor (~96m^3/d @ 50°→60°C) the temperature showed a gradual increase and by day 40 the digester temperature had risen to 38°C. With the temperature and volume of sludge from the aerobic reactor reduced (~24m^3/d @ 40°) between days 41-73 to allow for the modifications to be made to the recirculation line, the digester temperature showed a slight decline to 35°C. With the resumption of normal operation on day 74, the digester temperature started increasing again. By the end of the evaluation period (day 152) the temperature had stabilised at around 53°C. This was caused by the high sludge flow rate and temperature from the aerobic reactor to the anaerobic digester during the latter phase of the evaluation period (192 m^3/d @ 60°→65°C). Therefore operating the Athlone dual digester at an aerobic reactor retention time of ~1 day with a subsequent anaerobic digester retention time of ~10 days, forced **the anaerobic digester to operate in the thermophilic temperature range**. Under these aerobic operating conditions a heat exchanger between the reactor influent and effluent sludge flows would be required to maintain the digester in the mesophilic temperature range.

7.2.5 Aerobic Reactor Retention Time

The average aerobic reactor retention time during phase II (excluding days 41-73) was 1.6 days. During the first 8 steady state periods (until day 117) the reactor was operated at an approximate retention time of 2.0 days . During steady state period 9 (days 118-129) the retention time was about 1.4 days and for the final two steady state periods 10 and 11 (days 130-152), the retention time was around 1.0 day. The variation in reactor retention time (as with reactor temperature) was imposed on the reactor as part of the research requirements (see Figure 7.1), and was not a consequence of inherent system instability.

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longer and the feeding therefore more continuous. The Milnerton reactor was fed on the better draw and fill basis and because the batch volume was a constant $^1/_{12}$ of the reactor volume, a saw-tooth temperature profile in response to the draw and fill conditions was obtained at steady state (Messenger *et al*, 1992). Draw and fill feeding is superior because it ensures that there is no mixing between influent and effluent sludge and therefore avoids recontamination of pasteurised sludge.

7.2.6 Anaerobic Digester Retention Time

The average anaerobic digester retention time during phase II (excluding days 41-73) was 15.8 days and varied between 9 and 20 days (Table 7.1). The retention times achieved in the digester were a direct result of the feed sludge flow rates imposed on the aerobic reactor to achieve the desired retention times in the aerobic reactor (see Section 7.2.5 above). Because higher volumetric loading rates were made possible during phase II with the injection of pure oxygen, the retention time in the anaerobic digester fell below 12 days during the latter part of the period (days 130 - 152) and operated at an approximate 9 day retention time. It was therefore possible to test the claim for the dual digestion system that aerobic pre-treatment reduces the minimum anaerobic digester retention time to achieve sludge stabilisation. The tests carried out at laboratory scale during phase I (see Section 4.5) could now be verified at full scale.

7.2.7 Change in Sludge Characteristics during Dual Digestion

On average during phase II, approximately 20% of the volatile solids in the feed sludge were removed (destroyed) by treatment in the aerobic reactor, with a further 39% removal in the anaerobic digester. The average overall percentage removal of volatile solids across the process was 52% (Tables 7.1 and 7.15). A full discussion on the system total solids, volatile solids, and COD removal (reduction) is given in Section 7.4.1 below.

The conditioning effects of aerobic treatment was noticeable by; (1) an increase in average primary sludge feed pH of 5.4 to 7.2 and (2) an increase in the average ammonia and bicarbonate alkalinity levels from 108 mg(N)/l and 180 mg($CaCO_3$)/l to 518 mg(N)/l and 850 mg($CaCO_3$)/l respectively after aerobic treatment. Aspects of aerobic conditioning are discussed in detail in Section 7.4.3 below.

7.3 Evaluation of Aerobic Reactor Performance

7.3.1 The Steady State Periods

As in phase I, phase II was divided into a number of distinct steady state periods of stable operation ,to allow the heat balance and oxygenation characteristics for the reactor to be determined. A total of 11 steady state periods were defined, which differed from one another with respect to;

- air flow rate to the aerobic reactor;
- pure oxygen supply rate to the aerobic reactor;
- feed sludge temperature and flow rate to the aerobic reactor;
- hydraulic flow rate and pressure in the sludge recirculation line;
- ambient temperature.

These differences are listed in Table 7.2 below and can be seen graphically in Figure 7.1. During each of the eleven periods all the controllable parameters were held as constant as practically possible to allow steady state conditions to develop. The average data for each steady state period were calculated on the last 7 days data of each particular period to allow establishment of steady state conditions.

Table 7.2 Average Values of the Aerobic Reactor Operating Parameters for the Eleven Steady State Periods of Operation During Phase II

No:	Period Dates	Period Days Nos	Sludge flow m^3/d	Aerobic SRT days	Feed Conc g/ℓ	Oxygen Flow kg/h	Air Flow m^3/h	Feed sludge temp °C	Aerobic sludge temp °C
1	94-07-21 to 94-07-30	6-15	96	1.92	50	60	260	18	50
2	94-07-31 to 94-08-09	16-25	96	1.92	49	60	610	19	53
3	94-08-10 to 94-08-16	26-32	96	1.92	51	60	610	18	55
4	94-08-17 to 94-08-24	33-40	95	1.94	42	60	0	18	51
5	94-09-27 to 94-10-06	74-83	96	1.92	49	24	0	19	54
6	94-10-07 to 94-10-19	84-96	95	1.94	50	24	380	20	58
7	94-10-20 to 94-10-30	97-107	96	1.92	54	24	680	21	59
8	94-10-31 to 94-11-09	108-117	96	1.92	48	36	0	20	62
9	94-11-10 to 94-11-21	118-129	128	1.44	39	60	320	21	64
10	94-11-22 to 94-11-30	130-138	192	0.96	42	96	360	23	63
11	94-12-01 to 94-12-14	139-152	192	0.96	41	76	420	25	60

7.3.2 Measured Oxygenation Characteristics

Expressions for the supply, transfer, and utilisation of oxygen from both the air stream and pure oxygen injection stream, were derived in Section 3.3 above, in terms of the measurable parameters:

$Q(AIR)_{in}$ = The volumetric air flow rate into the aerobic reactor (m^3(STP)/h)

$M(O_2)_{in}^{O2}$ = The mass flow rate of pure oxygen injected into the sludge recirculation line (kg(O_2)/h)

$\%(O_2)_{out}^{AIR+O2}$ = The oxygen concentration in the dry effluent vent gas during oxygen supplementation ($\%^v/_v$)

$\%(O_2)_{out}^{AIR}$ = The oxygen concentration in the dry effluent vent gas when oxygenation is with air alone ($\%^v/_v$) i.e. the pure oxygen injection stream is temporarily switched off.[3]

[3] In order to solve the gas component mass balance it was necessary to measure the oxygen concentration in the vent gas with oxygen supplementation being both on and off. The procedure for doing this is described in Section 2.4.11.

Solution of the oxygenation terms (listed below) requires knowledge of the carbon dioxide concentration in the vent gas. Unfortunately, during phase II a suitable instrument was no longer available which would have allowed the vent gas carbon dioxide concentration to be measured. However, sufficient confidence was placed in the value determined for the respiration coefficient Y_{CO2} during phase I (see Section 4.3.6) for the value Y_{CO2}=0.70 mol(CO_2)/mol(O_2) to be used to allow for the solution of the oxygenation terms[4].

The equations derived in Sections 3.2 and 3.3 for the oxygenation terms are listed below. It was accepted that the pure oxygen injection system and the diffused air aeration system operate independently of each other; i.e. neither one influences the performance of the other, and that each system contributes individually to the overall oxygen utilisation rate; i.e. under oxygen limiting conditions:

$$OUR^{AIR+O2} = OUR^{O2} + OUR^{AIR} = OTR^{AIR+O2} = OTR^{O2} + OTR^{AIR} \quad ...kg(O_2)/m^3.h \quad (3.1)$$

The validity of this assumption was confirmed by the close correlation of the specific heat yield obtained during phase II compared with phase I (see Section 7.3.7), plus the observation that the pure oxygen transfer efficiency did not change with changing air flow rate (see Section 7.3.4). The individual equations are as follows:

The "Air" Oxygen Supply Rate (Eq 3.47)

$$OSR^{AIR} = \frac{0.279 \times Q(AIR)_{in}}{184}$$

The "Pure" Oxygen Supply Rate (Eq 3.92)

$$OSR^{O2} = \frac{M(O_2)_{in}^{O2}}{184}$$

The "Overall" Oxygen Supply Rate (Eq 3.93)

$$OSR^{AIR+O2} = \frac{0.279 \times Q(AIR)_{in} + M(O_2)_{in}^{O2}}{184}$$

The "Air" Oxygen Utilisation Rate (Eq 3.44)

$$OUR = \frac{Q(AIR)_{in}}{138} \cdot \frac{(21 - \%(O_2)_{out})}{(100 - 0.3 \times \%(O_2)_{out})}$$

---

[4] The average value for Y_{CO2} of 0.70 compares very favourably with the 0.67 value observed by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor.

The "Pure" Oxygen Utilisation Rate (Eq 3.88)

$$OUR^{O2} = \frac{M(O_2)_{in}^{O2}}{184} - \frac{\%(O_2)_{out}^{AIR+O2}\left(1.25Q(AIR)_{in} + 0.7M(O_2)_{in}^{O2}\right)}{184\left(100 - 0.3\%(O_2)_{out}^{AIR+O2}\right)} + \frac{1.25\%(O_2)_{out}^{AIR}.Q(AIR)_{in}}{184\left(100 - 0.3\%(O_2)_{out}^{AIR}\right)}$$

The "Overall" Oxygen Utilisation Rate (Eq 3.91)

$$OUR^{AIR+O2} = \frac{M(O_2)_{in}^{O2}}{184} + \frac{0.279Q(AIR)_{in}}{184} - \frac{\%(O_2)_{out}^{AIR+O2}\left(1.25Q(AIR)_{in} + 0.7M(O_2)_{in}^{O2}\right)}{184\left(100 - 0.3\%(O_2)_{out}^{AIR+O2}\right)}$$

The "Air" Oxygen Transfer Efficiency (Eq 3.49)

$$OTE = \frac{(100 - 4.76 \times \%(O_2)_{out})}{(100 - 0.3 \times \%(O_2)_{out})}$$

The "Pure" Oxygen Transfer Efficiency (Eq 3.95)

$$OTE^{O2} = 100 - \frac{\%(O_2)_{out}^{AIR+O2}\left(125Q(AIR)_{in} + 70M(O_2)_{in}^{O2}\right)}{M(O_2)_{in}^{O2}.\left(100 - 0.3\%(O_2)_{out}^{AIR+O2}\right)} + \frac{125\%(O_2)_{out}^{AIR}.Q(AIR)_{in}}{M(O_2)_{in}^{O2}.\left(100 - 0.3\%(O_2)_{out}^{AIR}\right)}$$

The average oxygen supply rates OSR^{AIR}, OSR^{O2}, OSR^{AIR+O2}, oxygen utilisation rates OUR^{AIR}, OUR^{O2}, OUR^{AIR+O2}, and oxygen transfer efficiencies OTE^{AIR}, OTE^{O2}, calculated for each of the steady state periods are shown in Table 7.3. A graphical representation of the variation in the oxygenation parameters throughout phase II is given in Figure 7.2.

Table 7.3 The Average Influent Air Flow Rate, Pure Oxygen Flow Rate, Oxygen Supply Rates, Utilisation Rates, and Transfer Efficiencies for Each Steady State Period During Phase II.

No.	$Q(AIR)_{in}$ m^3(STP)/h	$M(O_2)_{in}^{O2}$ kgO/h	OSR^{AIR} kgO/m^3.h	OSR^{O2} kgO/m^3.h	OSR^{AIR+O2} kgO/m^3.h	OUR^{AIR} kgO/m^3.h	OUR^{O2} kgO/m^3.h	OUR^{AIR+O2} kgO/m^3.h	OTE^{AIR} %	OTE^{O2} %
1	260	60	0.395	0.326	0.721	0.047	0.151	0.198	11.8	46.3
2	610	60	0.919	0.326	1.245	0.113	0.156	0.269	12.3	47.9
3	610	60	0.925	0.326	1.252	0.108	0.178	0.286	11.7	54.6
4	0	60	0	0.326	0.326	0	0.180	0.180	0	55.2
5	0	24	0	0.130	0.130	0	0.108	0.108	0	83.0
6	380	24	0.572	0.130	0.702	0.068	0.108	0.176	11.8	83.0
7	680	24	1.025	0.130	1.155	0.121	0.109	0.230	11.8	83.5
8	0	36	0	0.196	0.196	0	0.162	0.162	0	82.6
9	320	60	0.483	0.326	0.809	0.058	0.266	0.324	12.1	81.5
10	360	96	0.537	0.522	1.059	0.063	0.435	0.498	11.7	83.3
11	420	76	0.628	0.413	1.041	0.076	0.341	0.417	12.1	82.6

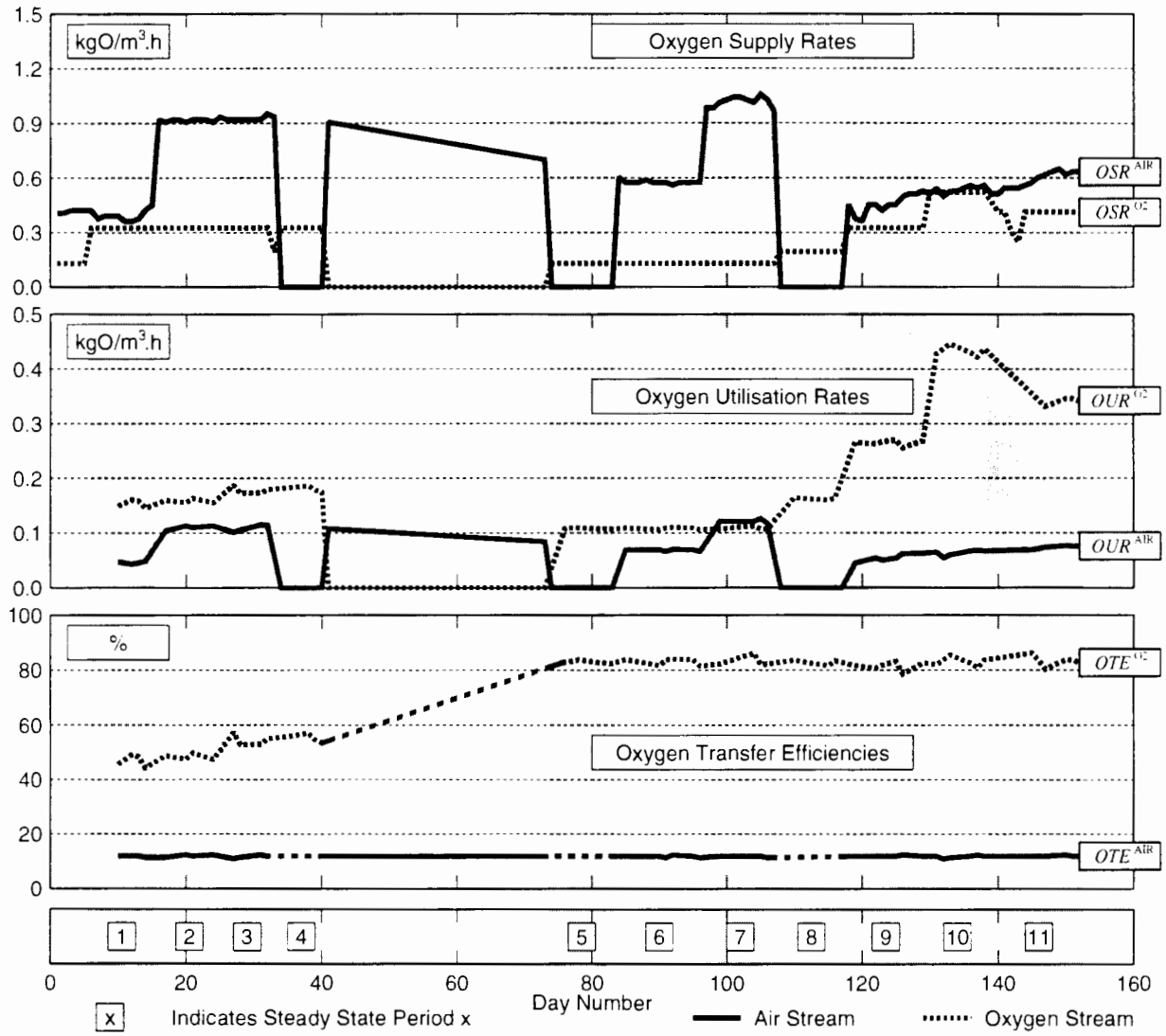

Figure 7.2 The Variation in the Oxygenation Characteristics during Phase II

Throughout the evaluation period there was no limitation on available substrate. This was verified by the fact that the dissolved oxygen level in the sludge was consistently less than 1 mg(O)/l. **The aerobic reactor was therefore oxygen limited throughout phase II**. This signifies that the rate at which the oxygen is utilised by the bacteria OUR^{AIR+O2} is fixed by the rates at which the oxygen is transferred into solution by the aeration device OTR^{AIR} and the pure oxygen injection device OTR^{O2}. The variation in the overall oxygen utilisation rate OUR^{AIR+O2} recorded during phase II is depicted in Figure 7.3 below. The maximum recorded level for OUR^{AIR+O2} of 0.50 kg(O_2)/m^3.h occurred during steady state period 10. This implies that the maximum possible level for OUR_{max}^{AIR+O2} (the point at which the reactor becomes substrate limited) lies in excess of this

figure[5]. At Milnerton, which treated a mixture of primary and humus tank sludge, Messenger *et al* (1992) observed a maximum sludge OUR_{max} of around 0.38kg(O_2)/m^3.h for a feed solids concentration of 30g(VS)/m^3. The Athlone sludge is essentially primary sludge (the humus sludge fraction is estimated to be less than 5%) and more concentrated (38g(VS)/m^3) so a higher OUR_{max} is not unexpected, even though the link between OUR_{max} and sludge type and feed VS concentration is not simple. A full discussion on maximum oxygen utilisation rates and required feed sludge solids concentrations to prevent substrate limitation occurring under differing conditions is presented in Chapter 8 below.

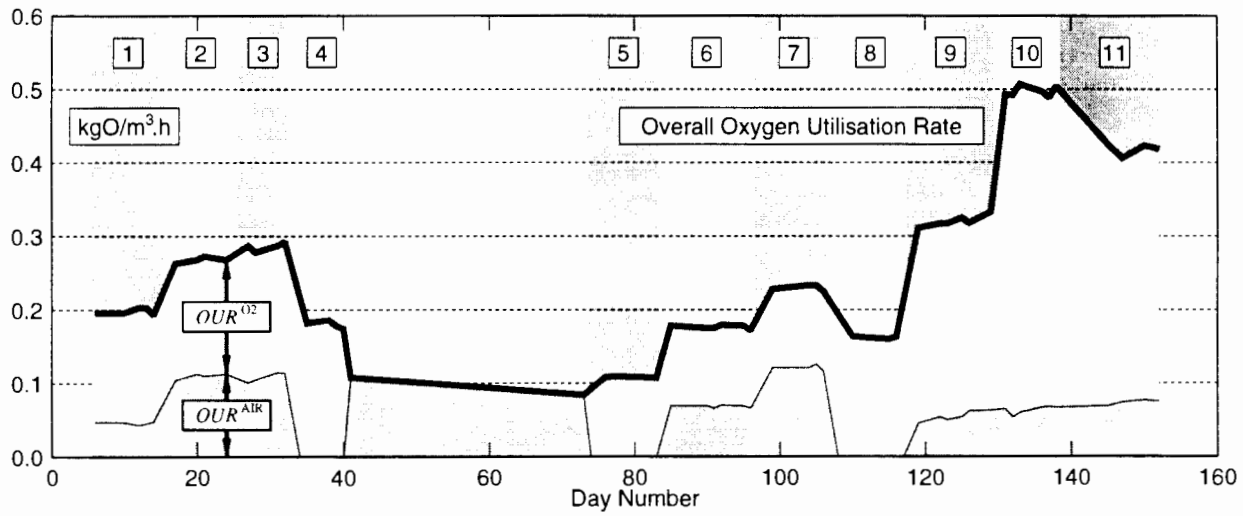

Figure 7.3 The Variation in the Overall Oxygen Utilisation Rate OUR^{AIR+O2} During Phase II

7.3.3 The Effect of Recirculation Flow Rate on Pure Oxygen Transfer Efficiency

The oxygen transfer efficiency OTE^{O2} which can be effected by the pure oxygen injection device is largely dependent upon the physical conditions within the sludge recirculation line. The principal effects are as follows:

- **The temperature, pressure and flow rate in the sludge recirculation line** which determines the maximum rate at which oxygen can be dissolved into the sludge after injection.

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[5] This figure is dependent upon the reactor retention time and the feed sludge solids concentration. During period 10 the retention time was at 0.96 days with a feed sludge solids of 42 kg/m^3 (see Table 7.2 above).

- **The velocity of the sludge through the Venturi** which affects the efficiency of mixing between the sludge liquid and gaseous oxygen.

- **The velocity of the sludge at the point of discharge back into the reactor** which affects the size of bubble formed as oxygen is released from solution due to the decrease in pressure upon entry into the reactor (the finer the bubble size, the greater the efficiency of re-dissolution of oxygen).

For the proposed incorporation of pure oxygen injection during phase II, the sludge recirculation line was modified (see Figure 10a.1 in Appendix 10). The motors and impellers on the pumps were changed to try and obtain the correct pumping conditions for successful pure oxygen injection; it was considered that a sludge flow rate of 700m^3/h, at a system head of 20m, were required. Subsequently, two further modifications were made to the pumping set-up during phase II to try and achieve the correct pumping conditions. The three sets of pumping conditions which existed during phase II are summarized in Table 7.4 below (see Appendix 10 for details).

Table 7.4 The Three Sets of Pumping Conditions in Operation During Phase II

Operating Period (Day No's)	No. of Pumps	Operational Configuration	Pump 1		Pump 2	
			Frequency (RPM)	Impeller Size (mm)	Frequency (RPM)	Impeller Size (mm)
1-25	1	-	1470	280	-	-
26-40	2	Parallel	1470	280	1470	280
74-152	2	Series	1617	350	1470	330

The flow characteristics in the recirculation line, in terms of flow rate, velocity at discharge and pressure for each pumping condition were determined in Appendix 10 below.[6] The results are tabulated in Table 7.5 below, along with the average pure oxygen transfer efficiency OTE^{O2} measured during each pumping period.

[6] This was achieved by superimposing the relevant pump characteristic curves onto the system characteristic curve.

Table 7.5 Flow Characteristics in the Sludge Recirculation Line for each of the Three Pumping Conditions During Phase II and the Measured Pure Oxygen Transfer Efficiencies OTE^{O2}

Operating Period (Day No's)	Recirculation Sludge Flow Rate (m^3/h)	Velocity at the Point of Discharge (m/s)	Pressure Head Generated in the Line (m)	Pure Oxygen Transfer Efficiency (%)
1-25	370	7.1	6.9	47.6±1.7
26-40	520	9.9	13.8	54.9±1.6
74-152	700	13.4	23.4	82.8±1.5

Statistical data for the measured pure oxygen transfer efficiency OTE^{O2} during each of the three pumping conditions is presented in Table 7.6 below:

Table 7.6 Statistical Data for the Measure Pure Oxygen Transfer Efficiency During Each Pumping Condition.

Recirculation Sludge Flow Rate (m^3/h)	Pure Oxygen Transfer Efficiency OTE^{O2} (%)				
	N° of Tests	Average	Std Deviation	Maximum	Minimum
370	8	**47.6**	1.7	49.8	44.3
520	8	**54.9**	1.6	57.3	52.8
700	36	**82.8**	1.5	86.2	78.6

During the third pumping condition (when two pumps were operated in series), the 75kW motor attached to the second pump was fitted with a variable speed drive which allowed the recirculation flow rate contributed by the second pump to be varied. On day 139 (1st December 1994) the pump motor was operated for 4×2 hour long periods at 35, 45, 55, and 60Hz frequency respectively. At the end of each 2 hour period the pure oxygen transfer efficiency OTE^{O2} was measured. The flow characteristics for each pumping condition were determined in Appendix 10 below. The results of these test are given in Table 7.7 below.

A summary table containing all the relevant measured and estimated data relating to the pumping, flow and oxygenation characteristics of the system is presented at the end of Appendix 10 (Table 10a.4).

Table 7.7 **Flow Characteristics in the Recirculation Line, at Different Speeds of the Second Pump in Series (1029-1764rpm), and the Measured Pure Oxygen Transfer Efficiencies OTE^{O2} ; Test Performed on Day 139**

Power Frequency (Hz)	Pump Speed (RPM)	At Total Recirculation Sludge Flow Rate (pump1+pump2)			
		Recirculation Sludge Flow Rate (m^3/h)	Velocity at Discharge (m/s)	Pressure Head Generated in the Line (m)	Pure Oxygen Transfer Efficiency (%)
35	1029	**560**	10.7	15.5	**67.8**
45	1323	**630**	12.0	18.8	**71.1**
55	1617	**700**	13.4	23.4	**82.2**
60	1764	**740**	14.2	26.4	**85.2**

These results (Table 7.7) were combined with the pure oxygen transfer efficiency data (OTE^{O2}) obtained during the whole of phase II for each pumping condition (Table 7.6). The variation in OTE^{O2} with flow rate, pressure, and discharge velocity is presented in Figure 7.4 below:

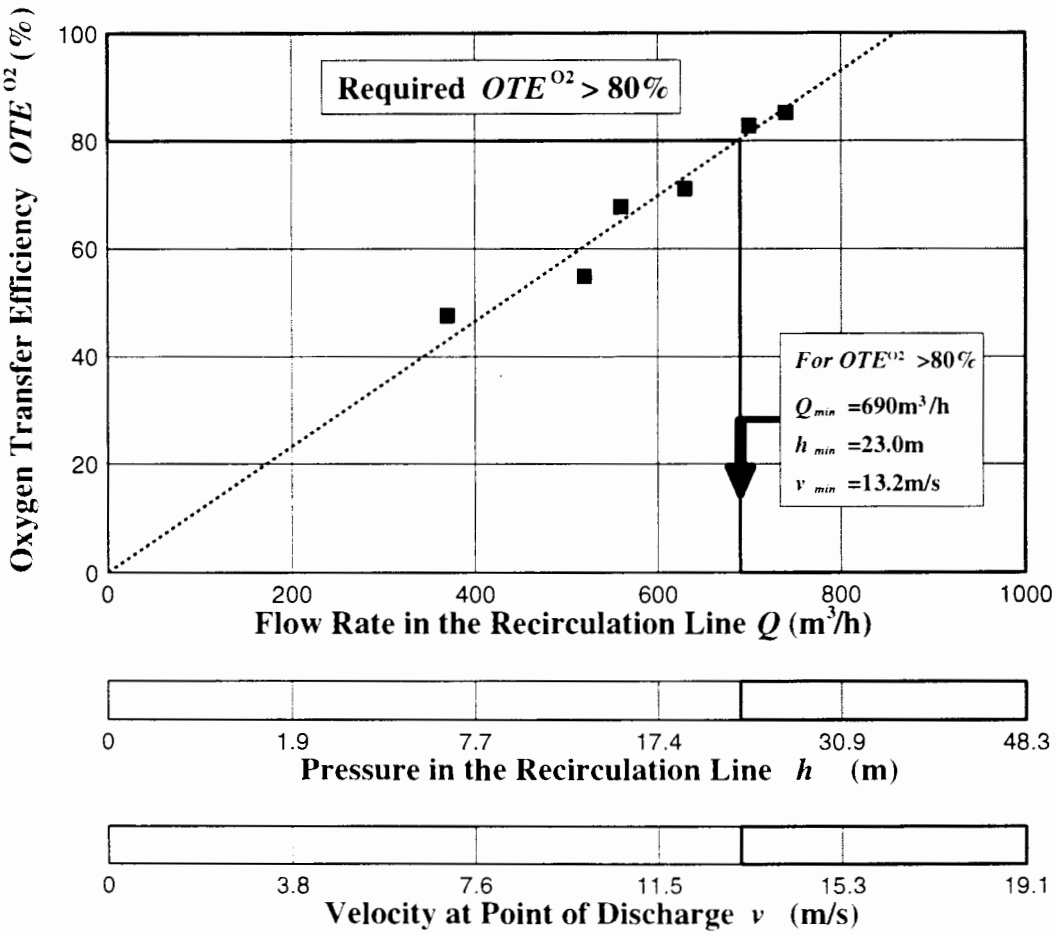

Figure 7.4 **The Variation in the Pure Oxygen Transfer Efficiency OTE^{O2} with Sludge Recirculation Flow Rate, Pressure and Discharge Velocity.**

Prior to the start of the investigation into pure oxygen supplementation (phase II), it was considered that a pure oxygen transfer efficiency in excess of 80% was desirable. From Figure 7.4 above, the flow characteristics necessary to achieve this (for conditions at Athlone) are presented in Table 7.8 below:

Table 7.8 Minimum Required Flow Characteristics in the Sludge Recirculation Line to Achieve a Pure Oxygen Transfer Efficiency OTE^{O2} of 80%

Minimum Required Recirculation Sludge Flow Rate (m^3/h)	Velocity at the Point of Discharge (m/s)	Pressure Head Generated in the Recirculation Line (m)	Minimum Required Pure Oxygen Transfer Efficiency (%)
690	13.2	23.0	**80.0**

7.3.4 The Effect of the Air Stream on the Pure Oxygen Transfer Efficiency

The effect of the air stream on the pure oxygen transfer efficiency was tested during the three steady state periods (N°s 5,6, and 7) following the establishment of the third pumping condition (i.e. two pumps operated in series). All operating conditions were held constant for these three steady state periods except the air flow rate (see Table 7.2; or Table 7.9 below).

Table 7.9 Aerobic Reactor Operating Parameters During Steady State Periods 5,6, and 7 of Phase II. All Parameters Held Constant Except Air Flow Rate. The Average Recorded Pure Oxygen Transfer Efficiency for each Period is Quoted.

No:	Sludge flow m^3/d	Aerobic SRT days	Feed Conc g/ℓ	Oxygen Flow kg/h	Air Flow m^3/h	Feed sludge temp °C	Aerobic sludge temp °C	Pure Oxygen Transfer Efficiency %
5	96	1.92	49	24	**0**	19	54	**83.0**
6	95	1.94	50	24	**380**	20	58	**83.0**
7	96	1.92	54	24	**680**	21	59	**83.5**

From the recorded average values of the pure oxygen transfer efficiency during steady state periods 5,6, and 7, it can be concluded that "**The Air Stream does not exert any influence on the Pure Oxygen Transfer Efficiency**".

7.3.5 The Effect of Pure Oxygen Injection on the Air Stream Oxygen Transfer Efficiency

Throughout phase II, regardless of the rate of pure oxygen injection into the sludge recirculation line, the oxygen transfer efficiency of oxygen from the air stream OTE^{AIR} did not vary significantly from an average efficiency of 11.9% (Table 7.10).

Table 7.10 Statistical Data for the Measured Oxygen Transfer Efficiency of Oxygen from the Air Stream; Throughout Phase II

Air Stream Oxygen Transfer Efficiency OTE^{AIR} (%)				
N° of Tests	Average	Std Deviation	Maximum	Minimum
39	11.9	0.38	12.4	10.9

The consistency in the OTE^{AIR} level throughout phase II is reflected in Figure 7.2 above (bottom line) and in the average values recorded for each steady state period (see Table 7.3 above). From this it can be concluded that "**Pure Oxygen Injection does not exert an influence on the Oxygen Transfer Efficiency of Oxygen from the Air Stream**".

Both conclusions, that of the influence of the air stream on the pure oxygen transfer efficiency and vica versa, validate the assumption made in Chapter 3 (derivations of solid, liquid and gas mass and heat balance equations - Section 3.3) that "**the air and pure oxygen streams can be considered to act independently of each other**".

7.3.6 The Steady State Heat Balance

As was the case during phase I, the heating performance of the aerobic reactor is assessed by establishing a steady state heat balance around the aerobic reactor. The main operating parameters are kept as constant as practically possible for a fixed period of time, thus allowing the reactor to reach steady state conditions. At steady state (manifested by constant temperature), the heat sources equal the heat sinks. The overall steady state balance (introduced in Section 3.1 and as employed in Chapter 4) is expressed as follows:

$$H_b + H_m = H_g + H_s + H_v + H_w \quad \text{...MJ/h} \quad (3.8)$$

where:

H_b = The rate of biological heat generation

H_m = The rate of heat energy input from the mixing device

H_s = The rate of sensible heat loss with the sludge leaving the reactor

H_v = The rate of vapour heat loss with the effluent gas stream

H_g = The rate of sensible heat loss with the effluent gas stream

H_w = The rate of heat loss through the walls of the reactor

During Phase I (when oxygenation was with air alone) the rate of biological heat generation H_b was found to be directly proportional to the oxygen utilisation rate OUR^{AIR}. The constant of proportionality Y_h, defined as the specific heat yield coefficient, was calculated to be 12.8 MJ/kg(O_2) (Section 4.3.8).[7]

The same procedure that was followed to calculate Y_h for phase I was used in phase II, i.e. with the exception of the biological heat generation rate H_b, each term in the steady state heat balance was calculated from the measurements made on the reactor. The biological heating rate was then calculated by difference with Eq 3.8 knowing the other heat loss and gain terms. A good comparison with the phase I determined value for Y_h would confirm the accuracy of the heat and mass balances of phase II and validate the acceptance of a value of 0.70 mol$(CO_2)_{gen}$/mol $(O_2)_{ut}$ for the respiration quotient Y_{CO2} (measured during phase I but assumed for phase II because a suitable instrument was no longer available to measure the vent gas carbon dioxide concentration).

Equations for the calculation of each of the heat balance terms (except H_b) when oxygenation is with air and pure oxygen were derived in Chapter 3 and are as follows:

The Water Vapour Heat Loss Rate in the Effluent Gas (Eq 3.133)
(Oxygenation with Air + Pure Oxygen)

$$H_v^{AIR+O2} = 2.38\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left(\frac{0.62\,a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{0.62\,a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}{1240-a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}\right)$$

The Effluent Gas Sensible Heat Loss Rate (Eq 3.153)
(Oxygenation with Air + Pure Oxygen)

$$H_g^{AIR+O2} = \left(1.21Q(AIR)_{in} + M(O_2)_{in}^{O2}\right).\left[0.001\left(T(AIR)_{out} - T(AIR)_{in}\right) + \frac{0.62\,a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}0.00187T(AIR)_{out}\right]$$

[7] This showed good agreement with the Y_h= 13.0 MJ/kg(O_2) determined by Messenger *et al* (1992) on the Milnerton pure oxygen aerobic reactor.

The Sludge Sensible Heat Loss Rate (Eq 3.163)
(Oxygenation with Air + Pure Oxygen)

$$H_s^{AIR+O2} = 4.04\left[\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{T(SL)_r\left(1.21\,Q(AIR)_{in} + M(O_2)_{in}^{O2}\right)}{1000}\cdot\frac{0.62 a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)}{760 - a\log_{10}\left(8.90 - \frac{2239}{273 + T(AIR)_{out}}\right)} - \frac{Q(SL)_{in}T(SL)_{in}}{24}\right]$$

THE Rate of Heat Loss from the Walls of the Reactor (Eq 3.195)
(Valid for Oxygenation with Air Alone and with Air + Pure Oxygen)

$$H_w = 0.82\left(T(SL)_r - T(SL)_d\right) + 0.49\left(T(SL)_r - T(AIR)_{in}\right)$$

The Rate of Mechanical Heat Input to the Reactor (Eq 3.200)
(Valid for Oxygenation with Air Alone and with Air + Pure Oxygen)

$$H_m = 1.65 \times I$$

The measurable parameters contained in the above formulae are;

$Q(SL)_{in}$ = The feed sludge flow rate (m^3/d)
$Q(AIR)_{in}$ = The dry air volumetric flow rate into the reactor (m^3(STP)/h)
$M(O_2)_{in}^{O2}$ = The pure oxygen mass flow rate into the reactor ($kg(O_2)/h$)
$T(AIR)_{in}$ = The influent gas stream temperature to the aerobic reactor (°C)
$T(AIR)_{out}$ = The effluent gas stream temperature from the aerobic reactor (°C)
$T(SL)_{in}$ = The temperature of the feed sludge to the reactor (°C)
$T(SL)_r$ = The temperature of the sludge leaving the reactor (°C)
$T(SL)_d$ = The temperature of the sludge in the digester (°C)
I = The current drawn by the mixing pumps (amperes)

The average values of the above monitoring parameters, obtained from the last seven days of each steady state period, are contained in Table 7.11 below:

Table 7.11 Average Steady State Values of the Monitoring Parameters Required for Solution of the Steady State Heat Balance: Phase II

No	Sludge Flow $Q(SL)_{in}$ m^3/d	Air Flow Rate $Q(AIR)_{in}$ m^3(STP)/h	Pure O_2 Flow $M(O_2)^{O2}_{in}$ kg(O_2)/h	Influent Air Temp $T(AIR)_{in}$ °C	Effluent Gas Temp $T(AIR)_{out}$ °C	Feed Sludge Temp $T(SL)_{in}$ °C	Reactor Sludge Temp $T(SL)_r$ °C	Digester Sludge Temp $T(SL)_r$ °C	Current Drawn by Pumps I Amps
1	95.8	261.5	60.0	15.6	47.0	17.8	50.2	34.2	75.9
2	96.2	608.6	60.0	17.3	51.0	19.2	53.5	36.3	73.6
3	95.9	612.9	60.0	16.3	52.8	18.3	55.5	37.0	96.0
4	95.5	0.0	60.0	17.9	46.2	18.2	50.6	37.2	94.3
5	96.0	0.0	24.0	17.5	50.0	18.9	54.0	37.5	204.9
6	95.3	378.6	24.0	19.3	55.2	19.9	57.9	42.8	208.8
7	95.8	678.6	24.0	20.9	56.0	21.0	58.6	47.3	207.3
8	96.0	0.0	36.0	19.0	57.4	20.5	61.9	50.0	205.6
9	127.9	320.0	60.0	21.3	60.8	21.3	63.6	51.2	205.3
10	192.0	355.8	96.0	23.2	60.2	23.3	63.3	52.9	205.3
11	191.0	415.8	76.0	23.8	57.5	24.6	60.3	53.5	203.6

From the operating data contained in Table 7.11 and the equations listed above, the heat balance terms for each steady state period were calculated and are listed in Table 7.12 below. A computer programme was compiled to perform the required calculations and this programme is listed in Appendix 8.

Table 7.12 The Calculated Heat Balance Terms for Each Steady State Period: Phase II Oxygenation with Air + Pure Oxygen

No	Heat Output Terms (MJ/h)				Heat Input (MJ/h)		Comment
	H_s	H_v	H_g	H_w	H_m	H_b	
1	512	60	14	30	125	**491**	O_2+AIR
2	534	160	34	32	121	**639**	O_2+AIR
3	575	180	37	34	158	**668**	O_2+AIR
4	515	9	2	27	156	**397**	O_2 ONLY
5	562	5	1	32	338	**262**	O_2 ONLY
6	590	124	23	31	344	**424**	O_2+AIR
7	575	226	40	28	342	**527**	O_2+AIR
8	662	11	2	31	339	**367**	O_2 ONLY
9	883	160	26	31	339	**761**	O_2+AIR
10	1259	181	29	28	339	**1158**	O_2+AIR
11	1123	168	28	24	336	**1007**	O_2+AIR

Note: The Biological Heating Rate H_b (shaded column) is calculated by difference.

7.3.7 Calculation of the Specific Heat Yield Coefficient

The biological heating rate H_b is directly proportional to the oxygen transfer rate *OTR*, which under oxygen limiting conditions, fixes the oxygen utilisation rate *OUR* (see Section 3.1). Oxygen limiting conditions were verified throughout phase II by the measurement of consistently low dissolved oxygen concentrations in the sludge liquid (≤1.0 mg(O_2)/l). The problem of computing the overall oxygen utilisation rate OUR^{AIR+O2} when two oxygen sources were involved (during phase II oxygenation was air + pure oxygen) was resolved by accepting that the two oxygen sources acted independently of each other. The validity of this assumption was verified in Sections 7.3.4 and 7.3.5 above. The biological heating rate is therefore given by:

$$H_b = Y_h.V_p.(OUR^{AIR} + OUR^{O2}) = Y_h.V_p.OUR^{AIR+O2} \quad \text{...MJ/h} \quad (3.201)$$

where:

Y_h = Specific heat yield coefficient in terms of oxygen utilisation (MJ/kg(O_2))

Vp = Aerobic reactor process volume (m^3)

OUR^{AIR} = Oxygen utilisation rate attributable to the oxygen from the **air** stream- volume specific (kg(O_2)/m^3.h)

OUR^{O2} = Oxygen utilisation rate attributable to the oxygen from the **pure oxygen** stream- volume specific (kg(O_2)/m^3.h)

OUR^{AIR+O2} = Overall oxygen utilisation rate attributable to the oxygen from both the **air** and **pure oxygen** streams- volume specific (kg(O_2)/m^3.h)

The biological heating rate H_b was determined by difference from the steady state heat balance (see Section 7.3.6) for each steady state period during phase II. The calculated values for H_b are given in the right hand column of Table 7.12 above. The overall oxygen utilisation rate OUR^{AIR+O2} was determined for each steady state period, by performing an oxygen mass balance (see Section 3.3) across the aerobic reactor. The calculated values for OUR^{AIR+O2} are given in Table 7.3 above (the separate contributions from the oxygen derived from the air stream OUR^{AIR} and the pure oxygen stream OUR^{O2} are also quoted). The linear relationship between the biological heating rate H_b and the oxygen utilisation rate OUR^{AIR+O2} is shown graphically in Figure 7.4 below. For comparison purposes the data obtained during phase I is also indicated on the graph (stars).

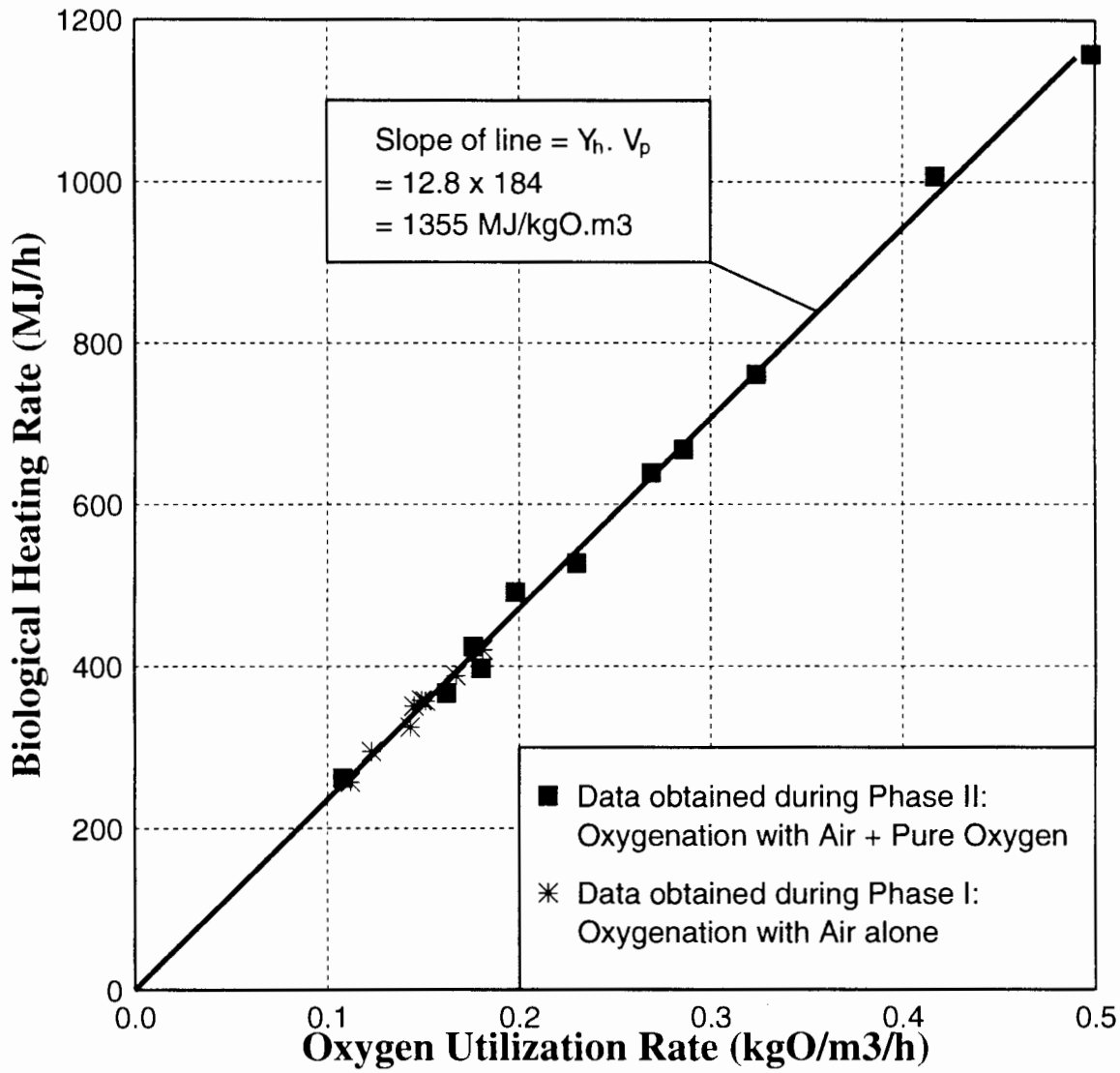

Figure 7.5 Biological Heating Rate versus Oxygen Utilisation Rate

The slope of the plot (from Eq 3.201) is equal to $Y_h \times V_p$. The calculated values for the specific heat yield Y_h for each steady state period are given in Table 7.13 below. Note that the aerobic reactor process volume V_p is equal to 184m^3.

Table 7.13 Calculated Specific Heat Yield Values for Each Steady State Period During Phase II: Oxygenation with Air + Pure Oxygen

No	OUR^{AIR+O2} $kg(O_2)/m^3.h$	H_b MJ/h	Y_h $MJ/kg(O_2)$
1	0.198	491	**13.5**
2	0.269	639	**12.9**
3	0.286	668	**12.7**
4	0.180	397	**12.0**
5	0.108	262	**13.1**
6	0.176	424	**13.1**
7	0.230	527	**12.5**
8	0.162	367	**12.3**
9	0.324	761	**12.8**
10	0.498	1158	**12.7**
11	0.417	1007	**13.1**

The calculated Y_h values vary in a very narrow band between 12.0 and 13.5 $MJ/kg(O_2)$ with the average value being 12.8 $MJ/kg(O_2)$. This value is in agreement with that determined during phase I (also 12.8 $MJ/kg(O_2)$) when oxygenation was with air alone, and compares favourably with the value of 13.0 $MJ/kg(O_2)$ obtained by Messenger *et al* (1992) on a pure oxygen aerobic reactor. This confirms the accuracy of the tests carried out at Athlone. The heat balance is proved to be reliably accurate under a wide range of operating conditions (see Table 7.1) both when oxygenation is with air alone and in combination with pure oxygen.

7.4 Detailed Aspects of Overall System Performance

7.4.1 System Volatile Solids and COD Reduction

Total Solids (TS), Volatile Solids (VS), and COD analyses were performed on a daily basis throughout phase II, excluding days 41 to 73 when the aerobic reactor sludge recirculation line was out of commission to allow for the pipework to be modified. The average TS, VS, and COD concentrations in the feed, aerobic, and anaerobic sludges for the evaluation period are given in Table 7.14 below.

Table 7.14 Average TS, VS, and COD Concentrations in the Feed, Aerobic, and Anaerobic Sludges During Phase II: Oxygenation with Air + Pure Oxygen.

Sludge Type	Total Solids (g/l)	Volatile Solids (g/l)	COD (g/l)
Feed	46.9	38.1	63.9
Aerobic	39.2	30.4	45.8
Anaerobic	27.4	18.4	28.8

A graphical representation of the variation in TS, VS, and COD concentrations in each stage of the process is presented in Figure 7.6 below. The data have been smoothed to eliminate short term fluctuations by the use of a weighted moving average filter prior to plotting.

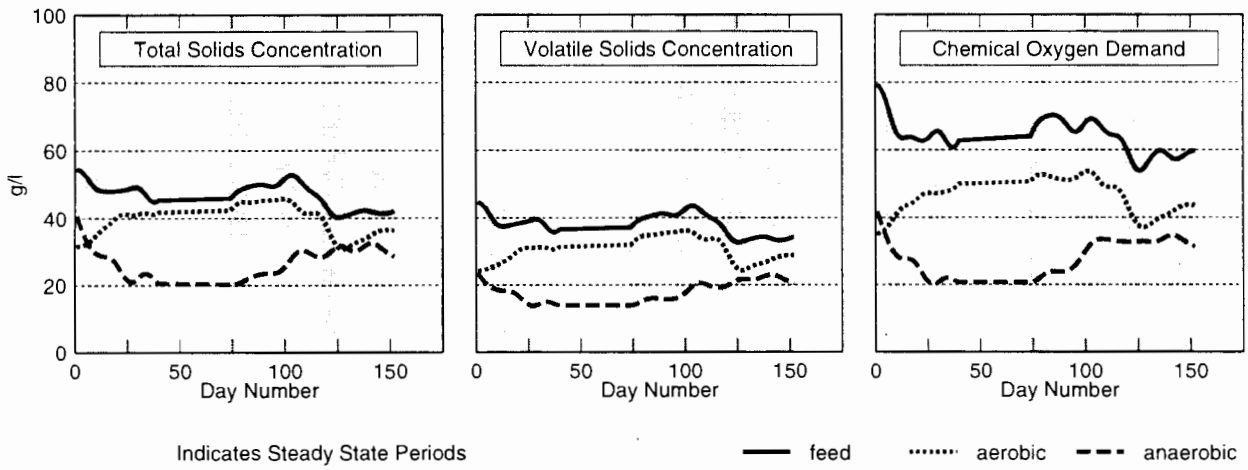

Figure 7.6 Graphical Representation of the Variation in Total Solids, Volatile Solids, and COD Concentrations for the Athlone Dual Digestion Process (Smoothed Data): Phase II Oxygenation with Air + Pure Oxygen

The variation in the percentage reduction in VS and COD mass across the aerobic reactor, anaerobic digester, and the overall process throughout phase II is shown in Figure 7.7. Again the data have been smoothed prior to plotting to eliminate short term variations. The rate of VS destruction in the aerobic reactor in relation to the oxygen utilisation rate is discussed in Section 7.5 below.

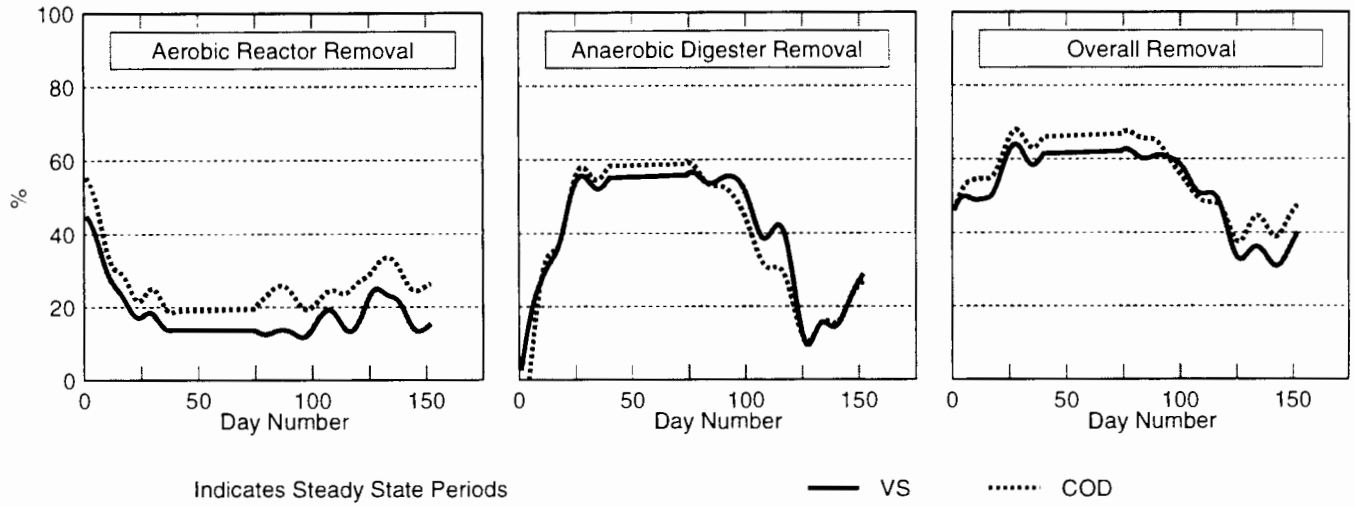

Figure 7.7 Graphical Representation of the Variation in Removal Efficiency for Volatile Solids, and COD for each stage in the Athlone Dual Digestion Process (Smoothed Data): Phase II Oxygenation with Air + Pure Oxygen

The average percentage removal efficiencies of TS, VS, and COD across the aerobic reactor, anaerobic digester, and the overall process are presented in Table 7.15 below. An average of 20% VS removal was achieved in the aerobic reactor. This is significantly higher in comparison with the 1.3% VS removal observed in the Milnerton pure oxygen reactor (Messenger *et al*, 1992). Further, Messenger *et al* (1992) found no relationship between VS reduction and oxygen utilisation. In contrast, for the Athlone aerobic reactor, a relationship between oxygen utilisation and volatile solids reduction could be established for retention times ranging from 1 to 5 days. The relationship was found to be closely equal to the COD/VS ratio of the raw feed sludge (see Section 7.5 below).

A significant decline in the removal efficiency of VS and COD across the anaerobic digestion stage is observed between days 100 and 140 (see Figure 7.7). A full explanation for this decline is given in Section 7.4.5 below.

Table 7.15 Percentage Reduction in Total Solids, Volatile Solids, and COD Across Each Stage in the Dual Digestion Process: Phase II Oxygenation with Air and Pure Oxygen

Treatment Stage	Total Solids (%)	Volatile Solids (%)	COD (%)
Aerobic	16.4	20.2	28.3
Anaerobic	30.1	39.5	37.1
Overall	41.6	51.7	54.9

7.4.2 Aerobic Conditioning

Aerobic conditioning was defined in Section 4.4.3 above as the effect of aerobic pre-treatment on the sludge which subsequently reduces the minimum required retention time in the anaerobic digestion stage. This is brought about mainly as a result of:

- an increase in alkalinity and pH as a result of ammonification and
- partial solublisation of particulate organic matter

An increase in pH and alkalinity as a result of ammonification after aerobic pre-treatment was manifest throughout phase II and the average values of the pH, conductivity, bicarbonate alkalinity and ammonia concentration of the feed and aerobic reactor sludges are given in Table 7.16 below; The smoothed variation in the pH, bicarbonate alkalinity, conductivity and ammonia concentrations of the feed, aerobic and anaerobic sludges throughout phase II is shown graphically in Figure 7.8.

Table 7.16 The Average pH, Conductivity, Ammonia, and Alkalinity Concentrations in the Feed, Aerobic and Anaerobic Sludges During Phase II

Sludge Type	pH		Conductivity mS/m		Ammonia Conc. mg(N)/ℓ		Bicarb Alk. mg($CaCO_3$)/ℓ	
	mean	range	mean	range	mean	range	mean	range
Raw Feed Sludge	5.4	4.8-6.2	266	131-386	108	17-200	180	0-460
Aerobic Sludge	7.2	6.7-7.9	450	161-675	518	82-826	850	280-1500
Anaerobic Sludge	7.2	6.7-7.6	609	239-790	753	480-1110	2870	1800-3870

Like in phase I, conditioning of the feed sludge through the partial solublisation of particulate organic matter in the aerobic reactor was not evaluated for phase II because of the difficulty of achieving solid/liquid separation of the aerobically treated sludges. Rather, the conditioning effects of aerobic treatment were assessed in terms of pH, conductivity, HCO_3^- ammonium and alkalinity increases. In this regard, Table 7.16 and Figure 7.8 show that the aerobic reactor sludge is a much more suitable feed for anaerobic digestion because it has a neutral pH, and high bicarbonate alkalinity. In phase I, it was found that the increases in ammonium, HCO_3^- alkalinity and pH were related to the VS reduction - the VS reduction, which increased with increased retention time, resulted in a proportional increase in ammonia, and because the increase in HCO_3^- alkalinity is principally due to ammonification (NH_3-NH_4^+), the VS reduction also resulted in a proportional increase in HCO_3^- alkalinity. The reduction in fermentation products in the primary sludge and the increase in HCO_3^- alkalinity in the aerobic reactor caused the

pH to increase into the 7-8 range. Comparison of the phase II conditioning effect results with those observed during phase I when oxygenation was with air alone (see Section 4.4.3) indicated that in phase II similar results were obtained. However, the link between the increase in ammonia concentration and HCO_3 alkalinity with VS removal is not as strong, possibly as a consequence of ammonia stripping.

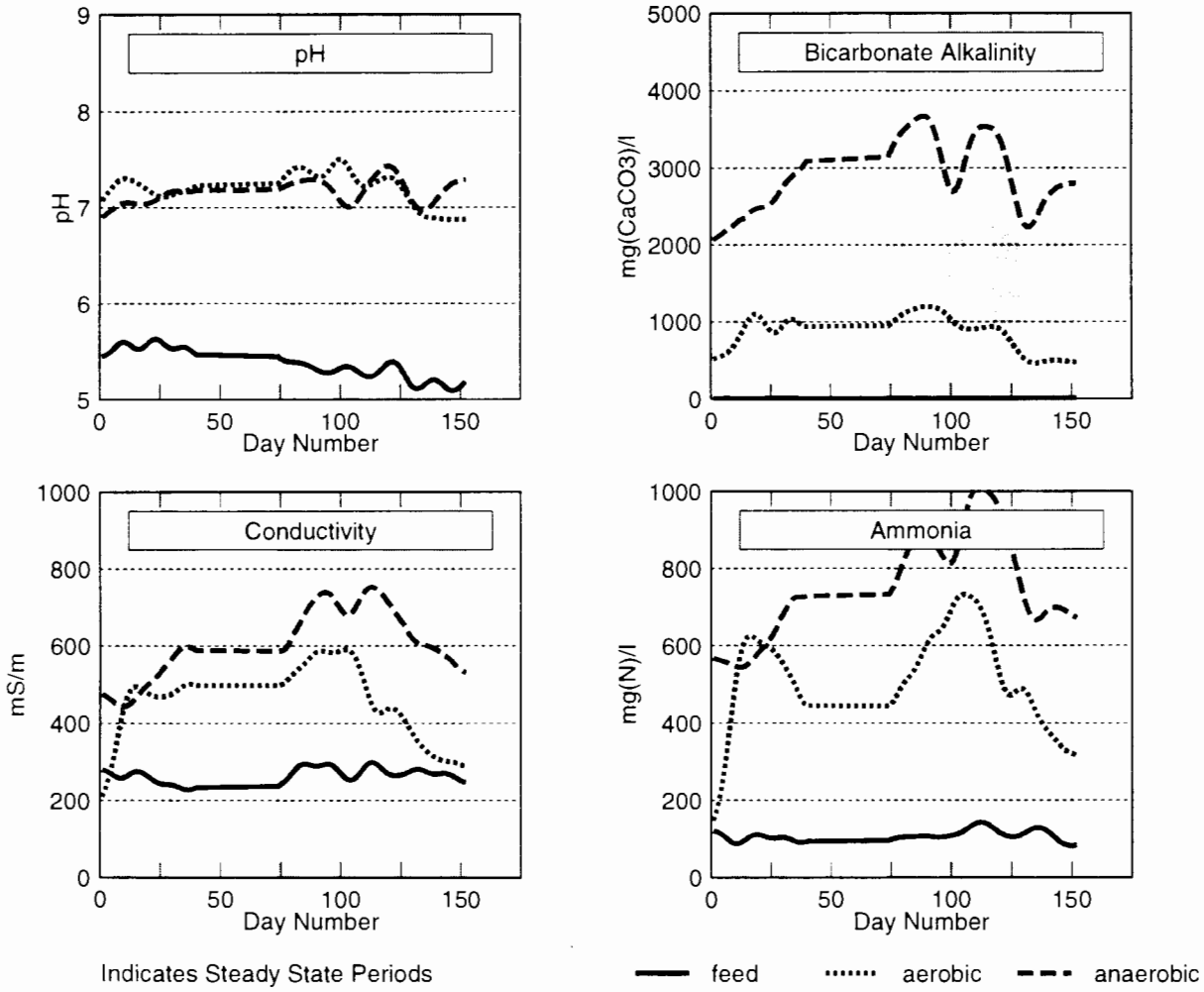

Figure 7.8 Graphical Representation of the Variation in pH, Alkalinity, Conductivity, and Ammonium Concentrations for the Athlone Dual Digestion Process (Smoothed Data): Results for Phase II

During phase I, the relatively low sludge flow rate through the process (mean rate 42 m^3/d) made it impossible to test (at full scale) the claim made for the dual digestion process that aerobic pre-treatment reduces the minimum required anaerobic retention time[8]. During phase II however, the use of pure oxygen made much higher sludge flow rates possible (maximum tested 192 m^3/d). With the anaerobic digester operating volume

[8] This claim was tested at the time by performing a laboratory scale study (see Section 4.5 above for details).

fixed at 1800m^3 , the shortest sludge retention time achieved during phase II was 9.4 days (ignoring 5-7% vaporisation water loss in the aerobic reactor). Consequently, it was of considerable interest to know how the digester performed at the high loading rates. However two factors influenced the anaerobic digester performance results, (1) the efficiency of gas mixing in the digester is notoriously poor (the biogas supply for mixing is low due to the other biogas requirements on the plant, namely the hot water heaters for the other primary anaerobic digesters and the gas engine) and (2) the high sensible heat content of the sludge from the aerobic reactor caused the digester temperature to increase into the thermophilic range. Earlier discussion on the level of volatile solids destruction in the digester (see Section 7.4.1) pointed out a decline in VS removal efficiency during the latter part of phase II, this was attributed to the transition of the digester from the mesophilic range into the thermophilic temperature range. Nevertheless, a thorough evaluation on digester stability and performance as well as sludge stability during phase II will give valuable insight into the claims made for dual digestion.

7.4.3 Anaerobic Digester Process and Product Stability

In view of the short retention times at which the anaerobic digester was operated during phase II (18 to 9 days), the stability of the anaerobic digestion *process* and of the sludge *product* were closely monitored. For the process this was done in terms of pH, bicarbonate alkalinity, and volatile acid alkalinity. For the final sludge product this was done in terms of %VS removal and specific oxygen utilisation rates (***SOUR***). It appears from the literature that the claim for dual digestion, *that aerobic pre-treatment reduces the minimum required anaerobic digester retention time,* has been focused on the anaerobic process stability with little cognizance of the stability of the final sludge product. It may well be that the digester operates satisfactorily at short retention times (~10 days) insofar as the anaerobic process is concerned but that longer retention times may be required because the final sludge product is (1) not stable and (2) has not been sufficiently reconstituted in the digester to dewater satisfactorily. These aspects are considered in this section.

It was pointed out earlier (Section 7.4.1), that between days 100 and 140, a significant decline in the digester VS removal efficiency took place (see Figure 7.7); the change in the principal monitoring digester parameters during this period is shown graphically in Figure 7.9 below.

The two factors which were of consequence to the operation of the digester during phase II were; (1) the high sensible heat content of the hot sludge from the aerobic reactor which forced the digester temperature into the thermophilic region (50°-70°C); and (2)

the relatively short retention time (reduced eventually to 9 days) which the digester operated under.

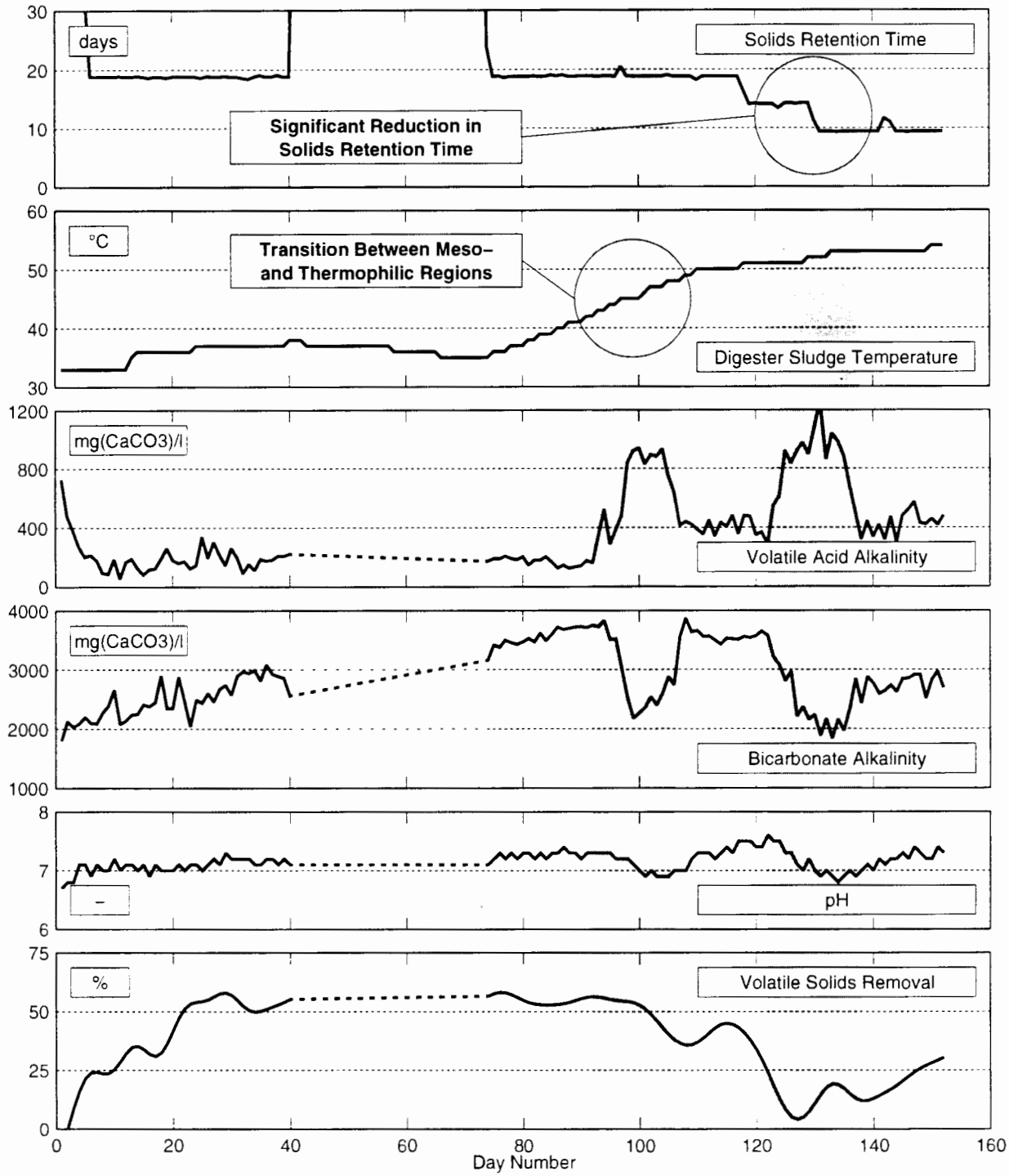

Figure 7.9 **Graphical Representation of the Variation in Solids Retention Time, Temperature, Volatile Acid Alkalinity, Bicarbonate Alkalinity, pH, and Volatile Solids Removal Efficiency (Smoothed) for the Anaerobic Digestion Stage During Phase II**

Examination of Figure 7.9 shows that the transition from mesophilic to thermophilic temperatures occurred between days 92-106. The reduction in solids retention time from

18 to 9 days (an increase in flow rate from $96m^3$ to $192m^3$) occurred in two steps at days 118 and 131. Both events caused the volatile acid alkalinity levels to temporarily increase to around 1000 $mg(CaCO_3)/l$ with a concomitant drop in pH to around 6.9. Sufficient buffer capacity was however available to absorb the effects (as indicated by the bicarbonate alkalinity levels) and in both instances normal digestion was resumed soon thereafter. The reduction in the efficiency of volatile solids removal is seen to have occurred during these periods, and this therefore offers an explanation for the decline. However, during the last 11^{th} steady state period (day 139-152; see Figure 7.7) the VS removal was showing significant increase again as the digester biomass was becoming accustomed to the thermophilic temperatures (~55°C).

In terms of the currently recognised measure of final sludge stability, a volatile solids removal efficiency of 51.7 was achieved by the overall dual digestion process (see Table 7.15). This is considered to be very good VS removal and well above that considered necessary for sludge stabilisation (>38%, Heidman, 1989).

Table 7.17 ***SOUR* Sludge Stability Test: Oxygen Utilisation (gO/kgTSS) During the First 24hrs of the Batch Test and the Average *SOUR*'s Recorded During this Period (gO/kgTSS.h) for the Different Sludge Types Tested**

Sludge Type	Oxygen Utilised During First 24hrs gO/kgTSS	Average *SOUR* During First 24hr Period gO/kgTSS.h
Dual Digested Sludge	81.6	3.4
- After Secondary Digestion	33.1	1.4
Waste Activated Sludge	46.6	1.9
- After Aerobic Digestion	15.6	0.7
Conventional Anaerobic Digested Sludge	86.4	3.6
Zimpro Sludge	68.6	2.9

The specific oxygen utilisation rate of the final sludge was also tested as a measure of sludge stability. A specific ***OUR*** of <1.0 $mg(O_2)/g(TS)$ is recognised as being indicative of a stable sludge (Heidman, 1989). The sludge from the dual digester at Athlone passes to a 2Ml secondary digester to enable dewatering to be carried out prior to discharge to the land drying beds. This sludge was also tested for specific ***OUR***, and for comparison purposes, a number of other different sludge types were tested. A full discussion on the specific ***OUR*** test method employed and the results obtained from the different sludge

types is presented in Appendix 7. The average specific ***OUR***'s recorded during the first 24 hours of the test for the different sludge types are tabulated in Table 7.17 above. The results of the tests are shown graphically in Figure 7a.2 (Appendix 7).

None of the sludges tested met with the 1g(O)/kgTSS.h target during the first 24 hours of the test except for the aerobically digested waste activated sludge. The stability of the sludge from the Dual Digester, in terms of the ***SOUR*** test, does however compare favourably with that observed for conventionally digested sludge. This inspite of the fact that the solids retention time through the dual digestion process is half that through the conventional process; a likely consequence of both stages in the dual digestion process operating at thermophilic temperatures (biological reactions proceed much faster at higher temperatures) and the conditioning effect of aerobic pre-treatment.

7.4.4 Bacteriological Quality

The degree of disinfection provided by the process during phase II was monitored by determining faecal coliform counts at each stage in the process[9]. The results obtained during the 152 day evaluation period are listed in Appendix 5. The median faecal coliform values obtained from this data are tabulated in Table 7.18 below.

Table 7.18 Summary of Faecal Coliform Analysis for Phase II

Faecal Coliforms/100mℓ - Medians		
Raw	Aerobic	Anaerobic
1.8×10^9	1.1×10^5	1.2×10^3

An approximate 4 orders of magnitude of reduction in faecal coliforms is observed across the aerobic reactor with a further 2 orders of magnitude reduction after anaerobic digestion. During phase I, when both aerobic reactor and anaerobic digester operated at lower temperatures, the reduction was lower i.e. 3 orders and 1 order of magnitude respectively (see Section 4.4.4 above).

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[9] It had been intended to monitor the destruction of viable *ascaris* ova through the process. However, analytical problems were encountered with the method (not clearly defined at this stage). Consequently, no data is available on the efficiency of destruction of viable *ascaris* ova

The population distribution of the values recorded for the feed, aerobic and anaerobic sludges during phase II are shown in Figure 7.10 below. Faecal coliform levels less than 10^3/100ml were reported as such (i.e. $<10^3$ per 100ml).

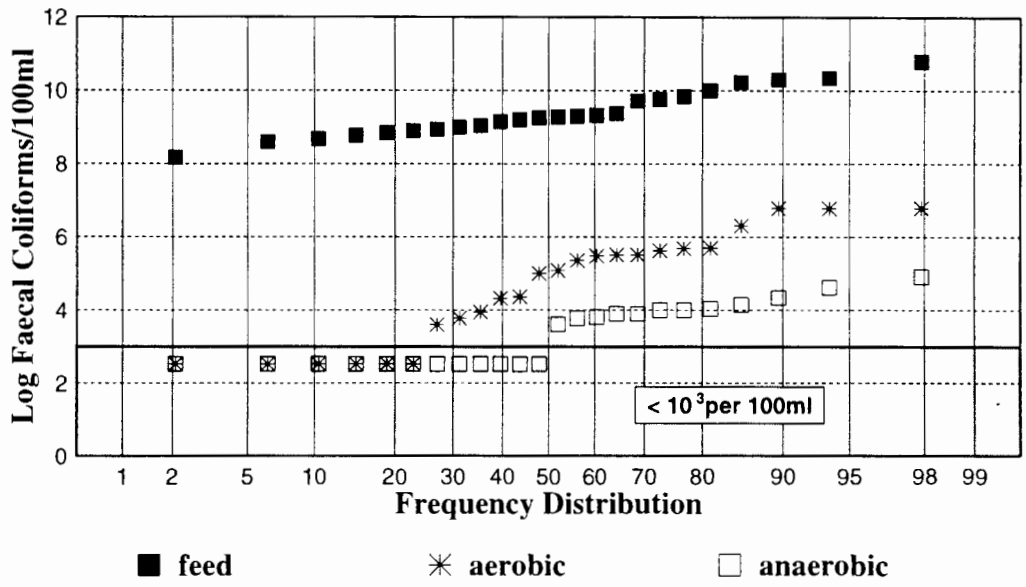

Figure 7.10 Frequency Distribution of Measured Faecal Coliform Counts: Phase II

The variation in faecal coliform concentration in the sludge from the **aerobic reactor** during phase II is shown in Figure 7.11 below. The reactor temperature and sludge flow rate are also indicated. Examination of Figure 7.11 shows that at a sludge flow rate of less than 100 m^3/d (2 day retention time) the faecal coliform concentration dropped below 10^3/100ml as the reactor temperature reached 58-59°C. However, as soon as the sludge flow rate was increased to 192 m^3/d (1 day retention time), the faecal coliform concentration in the sludge increased to between 10^4 and 10^6 /100ml. This would suggest that at the higher sludge loading rate, short circuiting is taking place in the aerobic reactor. This indicates that to provide effective pasteurisation, feeding on a draw and fill basis (as was done at Milnerton) would be required.

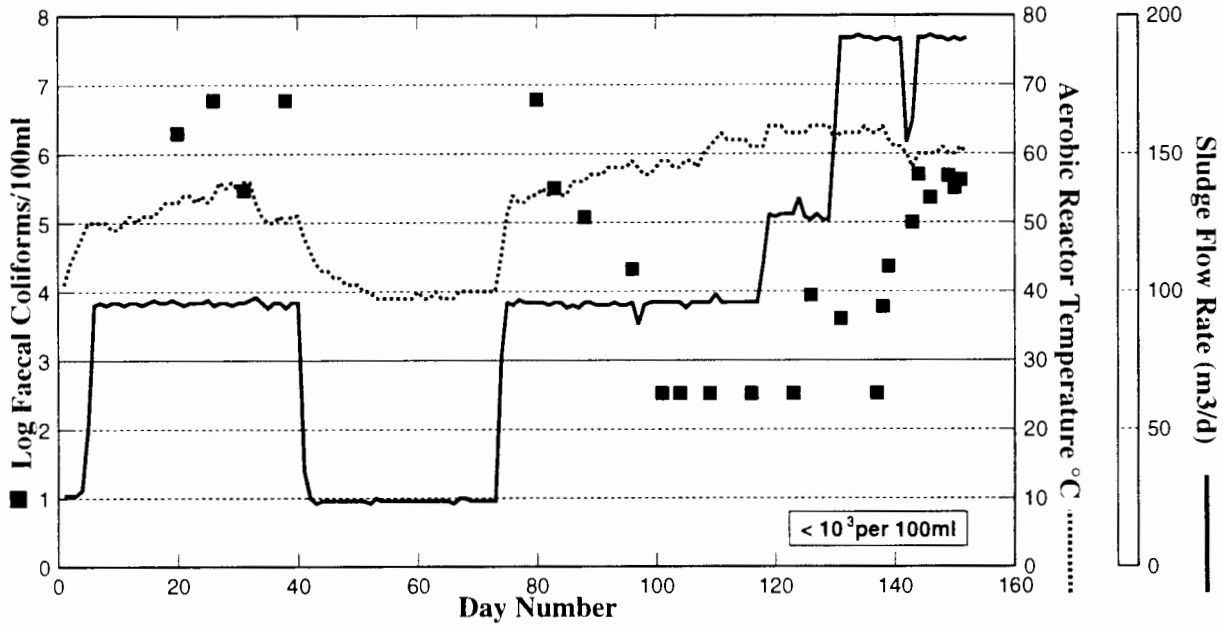

Figure 7.11 Graphical Representation of the Variation in Faecal Coliform Concentration (/100ml) in the Sludge Leaving the Aerobic Reactor During Phase II. Reactor Temperature and Sludge Flow Rate are Indicated

The variation in faecal coliform concentration in the sludge from the **anaerobic digester** during phase II is shown in Figure 7.12 below. The digester temperature and sludge flow rate are also indicated. Examination of Figure 7.16 shows that at a sludge flow rate of less than 100 m^3/d (18 day retention time) the faecal coliform concentration dropped below 10^3/100ml as the digester temperature entered the thermophilic region (50-70°C). It should be noted that at this time the faecal coliform concentration from the reactor was low (<10^3/100ml, see Figure 7.11) which could account for the low levels. Concentrations in excess of 10^3/100ml were observed at the higher sludge loading rate (9 day retention time), when higher concentrations of faecal coliforms (10^4- 10^6/100ml) were being discharged from the aerobic reactor.

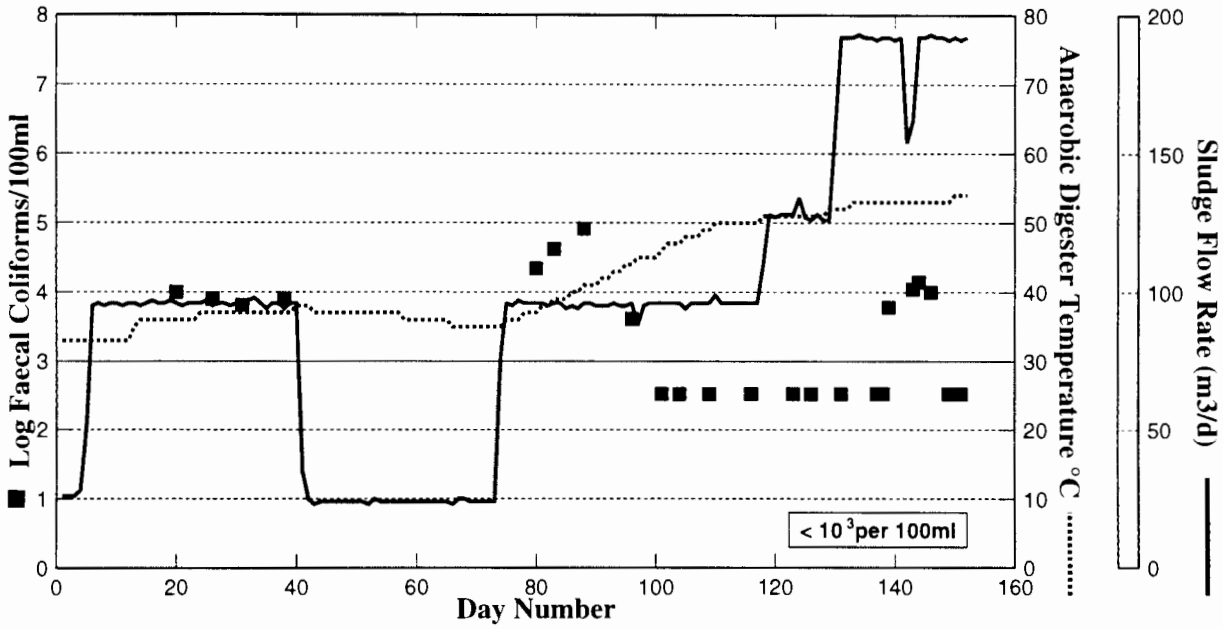

Figure 7.12 Graphical Representation of the Variation in Faecal Coliform Concentration (/100ml) in the Sludge Leaving the Anaerobic Digester During Phase II. Reactor Temperature and Sludge Flow Rate are Indicated

7.4.5 The Dewaterability of the Final Sludge

The dewaterability of the final treated sludge was measured by means of the specific resistance to filtration (SRF) test (Smollen, 1986 and Swanwick, 1962). A total of nine tests were performed during the latter part of phase II (between days 129 and 152), the results of which are listed in Appendix 5. These results are summarised in Table 7.21 below and comparison is made with values obtained after; (1) secondary digestion of the same dual digested sludge (see Section 7.4.6 for details); (2) Dual Digestion with Air - results obtained during phase I; (3) conventional mesophilic anaerobic digestion at the Athlone Wastewater Treatment Plant and (4) Dual Digestion with pure oxygen at Milnerton (Messenger *et al*, 1992).

Table 7.19 Comparison of the Dewaterability Characteristics of Sludges from the Dual Digestion Process, both with Air and with Oxygen, and the Anaerobic Digestion Process

Sludge Treatment	SRF (median) m/kg x 10^{12}	No of Tests	Sludge Temp Achieved
Dual Digestion using Air + Oxygen	802	9	65°C
Above after Secondary Digestion	237	4	-
Dual Digestion using Air	368	10	50°C
Dual Digestion using Oxygen	507	6	65°C
Mesophilic Anaerobic Digestion	218	10	37°C

By comparing the data in Table 7.19 above, it can be concluded that the dewaterability, in terms of the SRF test, of the different sludge types are not significantly different, and that all five sludges dewater rather poorly (c.f. Smollen, 1986).

After phase I, it was speculated that the temperature of treatment adversely affected the dewaterability (see Section 4.4.5). The results of the dewaterability studies performed during phase II, would tend to support this idea. The principal difference in operation between Athlone (phase II) and Milnerton (DD using oxygen), besides the use of air, is the fact that the Athlone anaerobic digestion stage operated at thermophilic temperatures. SRF values obtained for the sludge at Athlone during phase II are higher than those recorded at Milnerton (see Table 7.19 above). It is interesting to note, that after the sludge at Athlone underwent secondary digestion, the measured SRF values reduced to levels similar to that recorded for convention mesophilic digested sludges.

7.4.6 The Effect of Secondary Digestion

During phase II, the final sludge from the dual digester was passed to a secondary digester to allow for dewatering to take place. The secondary digester has a 2Mℓ capacity and is neither mixed nor heated. The unmixed (calm) hydraulic conditions in the tank allows for the sludge to settle and supernatant is withdrawn. Thickened sludge is normally drawn from the tank every few days (the frequency of which is dependent on the hydraulic loading rate) and pumped to land beds for drying. Figure 7.13 below shows the flow of sludge into and out of the digester (the difference between the two being the flow of supernatant). Because of the intermittent nature of desludging, smoothed data for the flow of sludge from the digester has also been plotted.

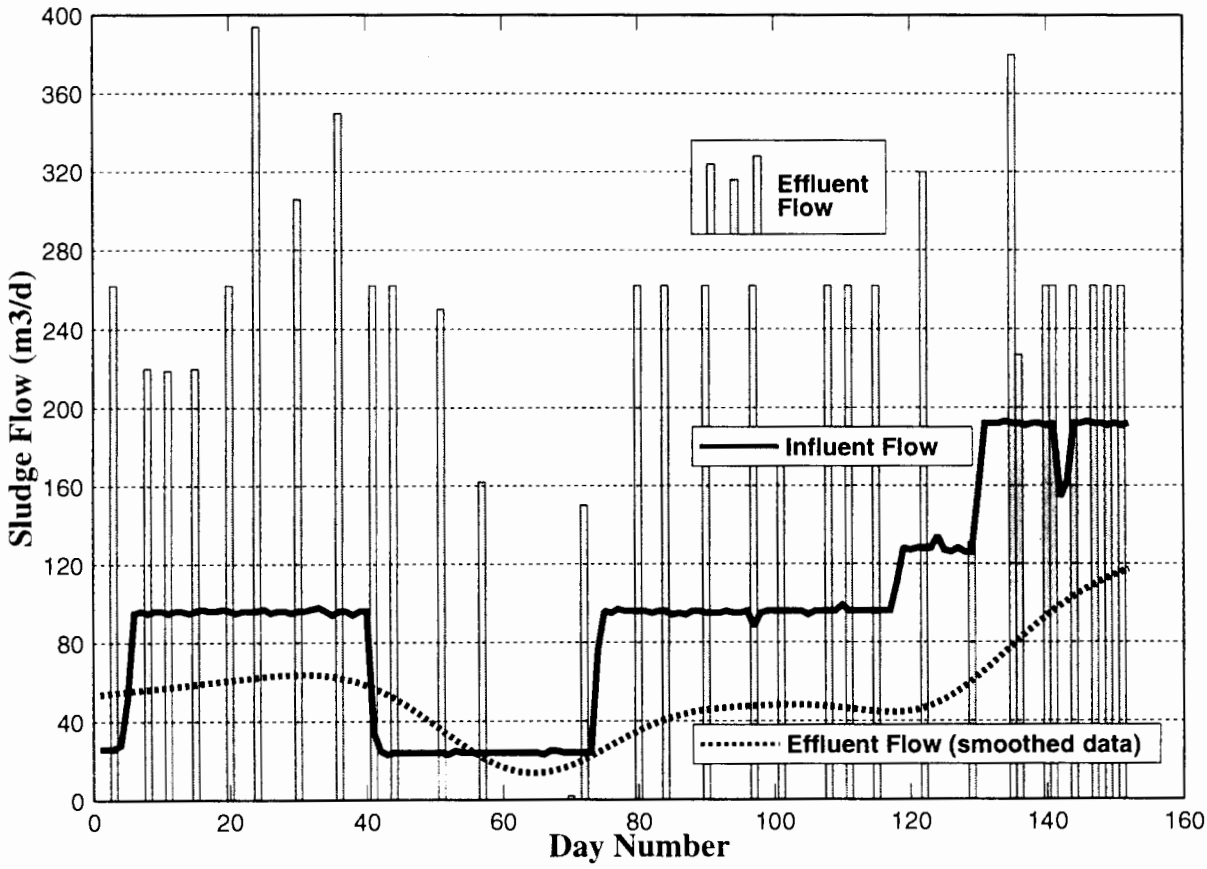

Figure 7.13 Variation in Influent and Effluent Flow Rate to the Secondary Digester Receiving Sludge from the Dual Digestion Plant: Phase II

The average flow rates into and from the secondary digester during phase II are listed in Table 7.20 below:

Table 7.20 Secondary Digester Sludge and Supernatant Flow Rates (m^3/d)

Sludge Type	Flow (m^3/d)
Influent Sludge (ex Dual Digester)	114
Effluent Sludge (to Drying Beds)	52
Supernatant Withdrawn (Recycled to ASP†)	62

† Activated Sludge Process

The sludge from the secondary digester was analysed weekly during phase II. The results of this analysis (mean, minimum, and maximum data) are presented in Table 7.21 below. These results are presented graphically in Figure 7.14 below.

Table 7.21 **Average Chemical Data for the Secondary Digester: Phase II**

Parameter		Average	Range
Temperature	°C	32	29-41
Total Solids	g/ℓ	53	20-69
Volatile Solids	g/ℓ	32	13-39
pH	-	7.2	7.0-7.5
Bicarbonate Alkalinity	mg($CaCO_3$)/ℓ	3700	2770-3700
Volatile Acid Alkalinity	mg($CaCO_3$/ℓ	140	0-540

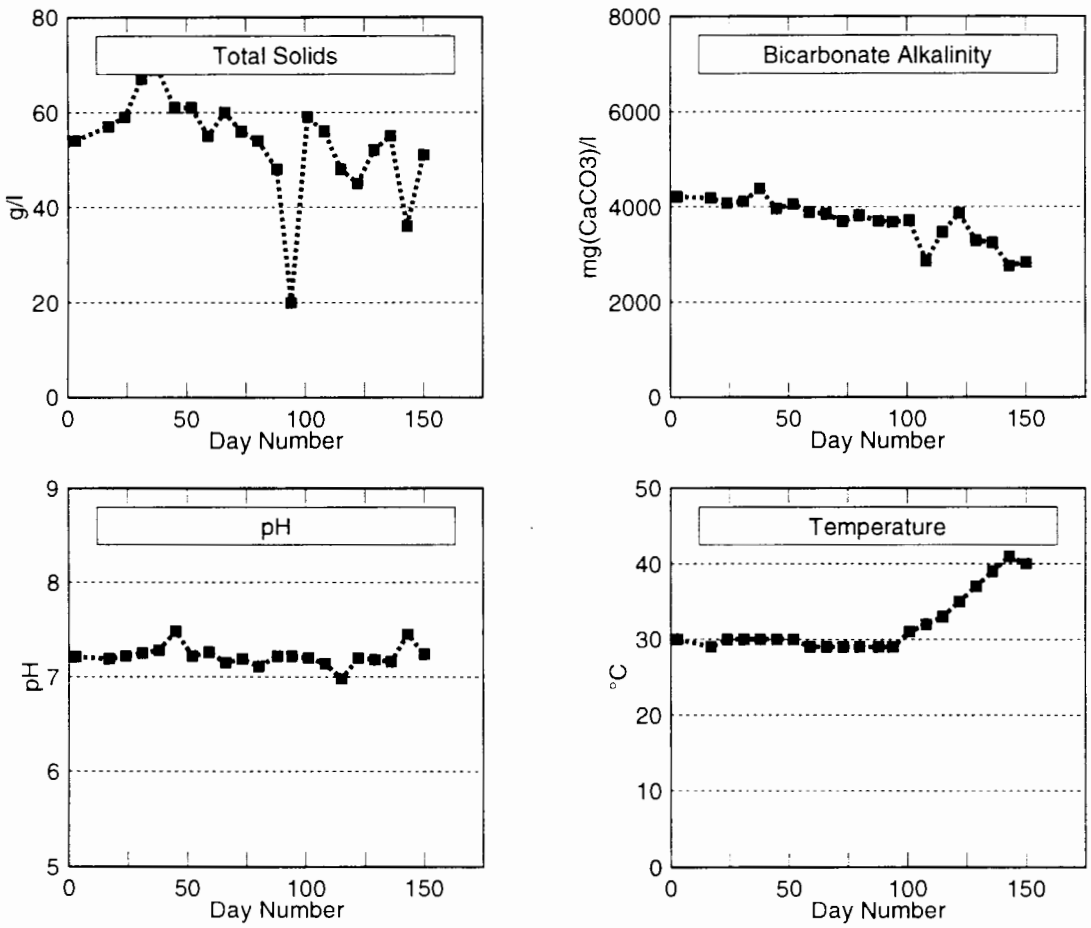

Figure 7.14 **Graphical Representation of the Variation in Total Solids, Alkalinity, pH and Temperature of the Sludge Leaving the Secondary Digester: Phase II. (This Digester Receives Sludge from the Dual Digestion Plant)**

Unfortunately, supernatant quality was not monitored during this investigation. Consequently, in order to assess the degree of volatile solids destruction taking place across the secondary digester, an estimate of the total and volatile solids concentration in the supernatant needs to be made. From an operating viewpoint, the operator makes a visual assessment of supernatant quality to determine draw-off times (quantity). In the past, samples of this supernatant have been analysed. The concentrations of Total Solids and Volatile Solids are relatively consistent at around 6.0 g(TS)/ℓ and 3.6g(VS)/ℓ respectively. Using this data to predict the volatile solids destruction yields the following (the volatile solids breakdown from the preceding stages are quoted for interest):

Table 7.22 Percentage Reduction in Volatile Solids Across Each Stage in the Dual Digestion Process, Including Secondary Digestion

Treatment Stage	Volatile Solids (%)
Aerobic	20.2
Anaerobic	39.5
Dual Digestion (Aerobic + Anaerobic)	51.7
Secondary Anaerobic Digestion	9.8
Overall (Dual + Secondary Digestion)	56.4

7.5 Relationship between Biological Heat, Oxygen Demand and VS Destruction

The relationship between the biological heating rate H_b and the oxygen utilisation rate *OUR* has been clearly shown (see Figure 7.5 above). At retention times ranging between 1.0 and 6.5 days (for operation with air, and air plus pure oxygen) the quantity of heat generated per mass of oxygen utilised (defined as the specific heat yield Y_h) was found to be closely constant at 12.8 MJ/Kg(O_2). Because this heat is generated through the oxidative degradation of the organics in the sludge, it is reasonable to expect that there will be a stoichiometric relationship between the rates of heat generation, oxygen utilisation and volatile solids destruction. Indeed, during phase I (when operation was with air alone, at an average retention time of 4.4 days) the biological heat generated per mass of VS destroyed was measured at 22 MJ/kg(VS) which showed good agreement with the 21 MJ/kg(VS) observed by Andrews and Kambhu (1971) for ATAD systems applications. Further, the mass of oxygen utilisation per mass of VS destroyed was

measured at 1.7 kg(O_2)/kg(VS) which is in close agreement with the COD/VS ratio for the feed sludge (see Section 4.4.2 above).

However, these findings do not concur with those observed by Messenger *et al* (1992) who found no such relationship(s) from their study on the pure oxygen reactor at Milnerton (operated at 1 to 3 day retention time). It could be considered that this discrepancy arises as a consequence of the difference in retention times: At Athlone, the longer retention times in operation during phase I, were more in accordance with those encountered in ATAD systems and therefore it is appropriate that the kinetics of VS destruction can be successfully applied. Messenger *et al* (1992) considered that the absence of a consistent relationship between VS destruction and oxygen utilisation could be attributed to the short retention times. They speculated that at very short retention times, the predominant biological activity was that of thermophilic organism growth as opposed to endogenous respiration, and for this reason ATAD VS destruction kinetics could not be applied. Operation of the Athlone aerobic reactor at short retention times (1 to 2 days), using air plus pure oxygen, allowed the relationship between heat generation, oxygen utilisation and VS destruction to be further examined.

Figure 7.15 below compares the day to day variation in the overall oxygen utilisation rate $M(O_2)_{ut}$ together with the calculated daily volatile solids destruction rate (smoothed) $M(VS)_{dest}$ during phase II. The similarity in the shapes of the curves gives strong indication that the two rate parameters are linked.

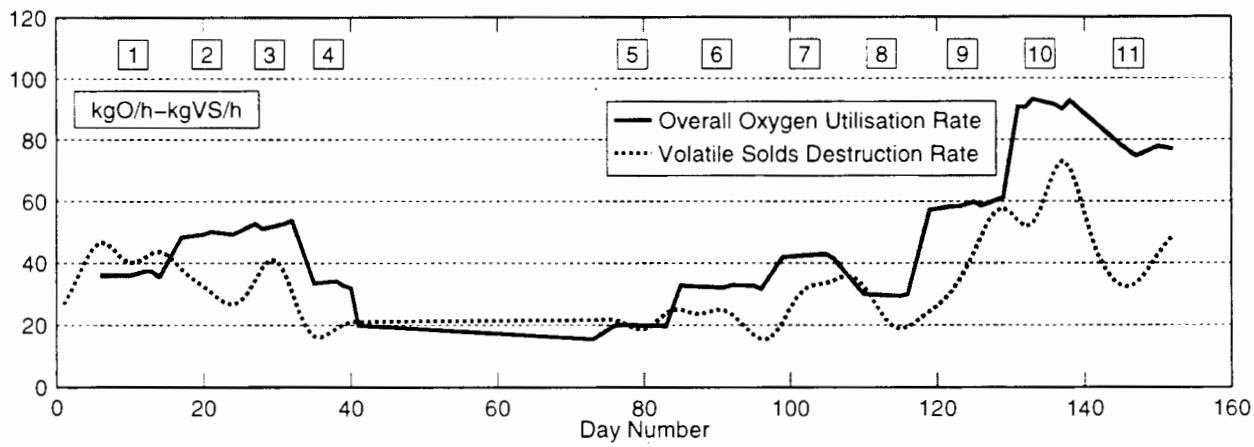

Figure 7.15 Graphical Representation of the Variation in the Overall Oxygen Utilisation Rate $M(O_2)_{ut}$ and the Volatile Solids Destruction Rate $M(VS)_{dest}$ in the Aerobic Reactor During Phase II: Oxygenation with Air + Pure Oxygen

The calculated average rates of biological heating, oxygen utilisation, and volatile solids destruction over prolonged periods of operation at Athlone (both phase I and phase II) are given in Table 7.23 below.[10] The three periods described differ principally in terms of retention time and the mode of oxygenation. The rates of VS and COD destruction were calculated, as described in Section 4.4.1 above, in accordance with Eq 4.03.

Table 7.23 The Average Rates of Biological Heating, Oxygen Utilisation and Volatile Solids Destruction for Differing Prolonged Periods of Aerobic Reactor Operation.

Phase (Day N°'s)	**Oxygen-ation**	R_h (d)	H_b (MJ/h)	$M(O_2)_{ut}$ (kg(O_2)/h)	$M(VS)_{dest}$ (kg(VS)/h)	$M(COD)_{dest}$ (kg(O_2)/h)	$\%(VS)_{dest}$
I (1-312)	Air Only	4.4	353	27.6	16.2	31.5	25
II (6-117)†	Air/O_2	1.9	438	34.3	25.1	62.5	16
II(118-152)	Air/O_2	1.1	961	75.7	47.0	118.1	20

† Excludes days 41-73 when the recirculation line was being modified

The relationship between the mass oxygen utilisation rate $M(O_2)_{ut}$ and the volatile solids destruction rate $M(VS)_{dest}$ is shown graphically in Figure 7.16 below. The ratio between the two parameters is approximately 1.6 kg(O_2)/kg(VS): indicated by the dotted line on the graph. This figure is in good agreement with the COD/VS ratio of 1.7 kg(O_2)/kg(VS) recorded for the raw feed sludge at Athlone.

[10] It is necessary to calculate the rate of VS and COD destruction over prolonged periods of time to reduce the influence of the variation in solids concentration in the feed sludge and the vagaries associated with taking representative samples.

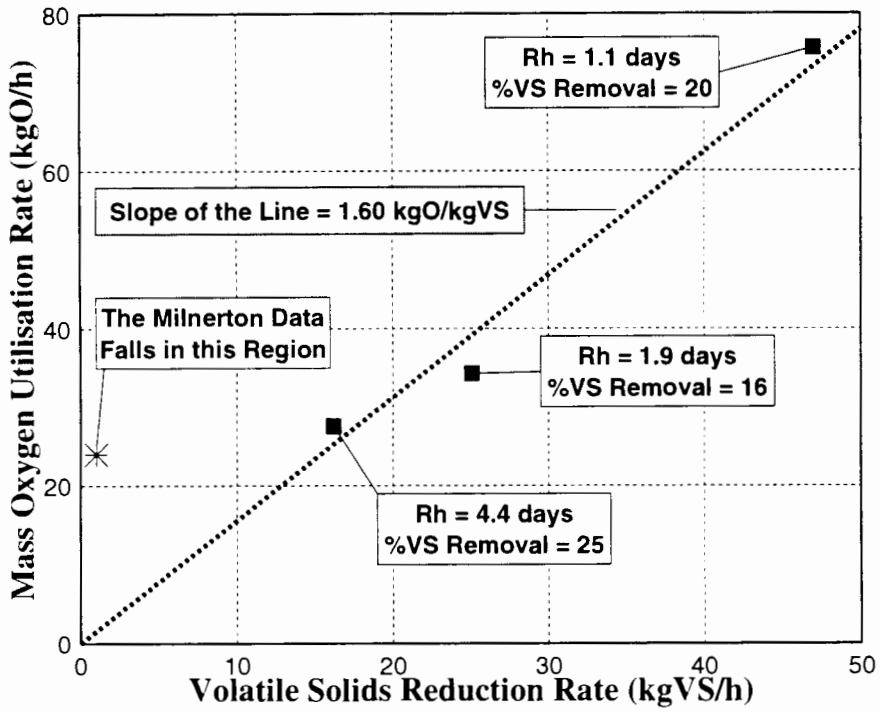

Figure 7.16 Graphical Demonstration of the Relationship Between Oxygen Utilisation Rate $M(O_2)_{ut}$ and the Rate of Volatile Solids Destruction $M(VS)_{dest}$ in the Aerobic Reactor at Different Retention Times.

From the data contained in Table 7.23 and presented graphically in Figure 7.16, a relationship between the rates of oxygen utilisation and volatile destruction is indicated. There is no intention to claim any direct linear relationship from the data (only 3 points), however the data does suggest that the relationship is close to the COD/VS ratio of the feed sludge, and as such can be used to predict the degree of VS removal obtained in the aerobic reactor. Whilst this contradicts the observation by Messenger *et al* (1992) who found little volatile solids destruction taking place in the aerobic reactor (<2%) at short retention times (1-2 days), it is in agreement with Appleton and Venosa (1986) and Hamer and Zwiefelhofer (1986) who recorded significant reductions under similar conditions (see Table 7.24). Oxygenation in all three dual digester aerobic reactors in

Table 7.24 Reported Volatile Solids Destruction (Percentage) Across the Aerobic Stage of Three Full Scale Dual Digestion Plants

Authors	Plants	Retention Time (d)	VS Removal (%)
Appleton and Venosa (1986)	Lackawana and Hagerstown, USA	1.1-2.0	10.8-19.9
Hamer and Zwiefelhofer (1986)	Untertezen, Switzerland	2.3	16.1

Table 7.24 was with pure oxygen. Unfortunately no data was provided with regard to mass oxygen utilisation and mass VS destruction rates for these plant, making a direct comparison difficult. However the degree of volatile solids destruction is similar to that observed at Athlone at the same retention time (1 to 2 days). It should be noted however (as apparent from Figure 7.16), that its VS destruction rate and not %VS removal that is consistently linked to mass oxygen utilisation rate.

The quantity of biological heat generated per mass of volatile solids destroyed $Y_{h\text{-}vs}$ during phase II averages at 19.0 MJ/kg(VS). The average figure measured during phase I was 22.0 MJ/kg(VS). Both figures show reasonable agreement with the 21.0 MJ/kg(VS) obtained by Andrews and Kambhu (1971) for ATAD systems applications.

The rate of COD destruction across the aerobic reactor $M(COD)_{dest}$ was consistently higher than the overall rate of oxygen utilisation $M(O_2)_{ut}$ during both phase I and phase II (see Table 7.16 above)[11]. During phase I (when oxygenation was with air alone) the rate of COD destruction was approximately 14% higher ($M(COD)_{dest}$=31.5kg(O_2)/h compared to $M(O_2)_{ut}$ = 27.6kg(O_2)/h) . During phase II, the rate of COD destruction was over 60% higher ($M(COD)_{dest}$ = 62.5 and 118.1kg(O_2)/h compared to $M(O_2)_{ut}$ = 34.3 and 75.7kg(O_2)/h). It is speculated that excess COD removal occurs as a result of enzyme hydrolysis; Particulate organics are solubilised by enzyme hydrolysis prior to being utilised by the aerobic bacteria. This action does not directly require the utilisation of oxygen. However, if the products of hydrolysis have a lower COD value than the initial organics, then a reduction in COD will be achieved without the utilisation of oxygen i.e. COD removal takes place without the transfer of electrons. It is possible that there could be an error in the COD measurement/sampling, however the VS results tend to conform to expectations indicating that the sampling techniques applied were adequate.

That a relationship exists between COD destruction and oxygen utilisation is evident however in Figure 7.17 below, which shows the variation in the overall oxygen utilisation rate $M(O_2)_{ut}$ together with the calculated daily COD destruction rate (smoothed) $M(COD)_{dest}$ during phase II. The similarity in the shapes of the curves gives strong indication that the two rate parameters are apparently associated.

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[11] Theoretically from a COD balance, the rate of COD destruction $M(COD)_{dest}$ should be equal to the overall rate of oxygen utilisation $M(O_2)_{ut}$.

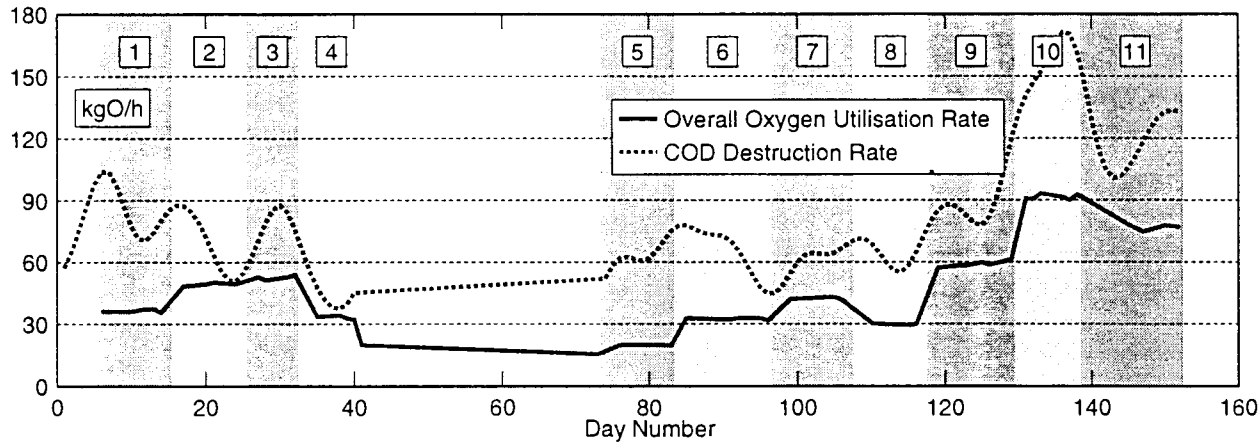

Figure 7.17 Graphical Representation of the Variation in the Overall Oxygen Utilisation Rate $M(O_2)_{ut}$ and the COD Destruction Rate $M(COD)_{dest}$ in the Aerobic Reactor During Phase II: Oxygenation with Air + Pure Oxygen

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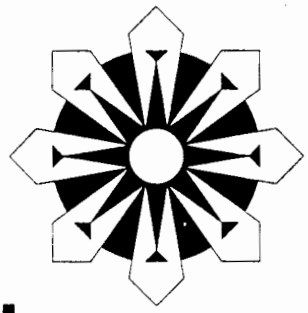


CHAPTER 8

A GENERAL SIMULATION MODEL FOR THE DUAL DIGESTION SYSTEM

8.1 INTRODUCTION

8.1.1 Description of the General Dual Digestion Model

As a result of this investigation (phase I and phase II), sufficient information has been obtained with regard to the dual digestion system to allow a mathematical model to be compiled which can simulate reasonable reliably all the main operating parameters in the system, and provide an assessment of system performance, both for the aerobic reactor and anaerobic digester, under a variety of different system configurations. Such a model was compiled as part of this investigation and is described in this Chapter.

The dual digester system model, which includes both aerobic reactor and anaerobic digester, is able to simulate aerobic reactor performance using air or pure oxygen alone or a combination of air and pure oxygen and a mesophilic or thermophilic anaerobic digester. The model has been constructed in such a way that it can be applied for any size and type of dual digester application, i.e. it is not site specific for Athlone. The principal parameters which are computed by the model are as follows:

- Oxygenation characteristics of the aerobic reactor
- Steady state heat balance for the aerobic reactor
- Heating requirements for the anaerobic digester
- Effect of installing heat exchangers
- Volatile solids destruction inside both the reactor and digester.
- Biogas production inside the digester
- Performance of an installed gas engine
- The stability of the final sludge.
- Minimum reactor retention time to prevent substrate limitation.
- Minimum digester retention time to ensure sludge stability.

The model includes a general cost analysis for both operating and capital costs. This allows approximate cost comparisons to be made between different system configurations, such as between the different dual digestion systems and conventional mesophilic anaerobic treatment both in terms of the parameters listed above (where applicable) and in terms of approximate capital and operating costs.

The model has been cast into a computer programme. The operation of the programme (Section 8.2), the basis and function of each of its components (Section 8.3) and the type of output produced (Section 8.4) are described in detail in this Chapter. A listing of the computer programme is given in Appendix 8. A number of examples of application of the model are presented in Section 8.5. These encompass:

1 The effect of ambient temperature on process performance
2 The effect of improved 'air' transfer efficiency on process performance
3 The effect of external reactor heating on process performance
4 The prediction of the onset of substrate limitation
5 The comparison between different system configurations

8.1.2 Running the Programme

The model has been compiled in the C programming language (see Appendix 8) and can be run on either the UNIX or DOS operating system. The programme is started by running the batch file **ddsim**. Upon running the programme, the user is prompted to answer a number of questions in order to define the dual digestion system to be simulated. Details regarding the information required by the programme is provided in Section 8.2 below. The programme then performs the necessary computations for each of the components of the model (details of which are presented in Section 8.3 below). Upon completion, the simulation results are given as screen output. The user is given the option to produce a hard copy of the results (an example is given in Section 8.4; Figure 8.8).

8.2 MODEL SET-UP

8.2.1 Introduction

The different input responses required to define a specific system configuration at the start of the simulation fall into the following categories:

- System retention times (dual digestion or conventional digestion)
- Size, number and shape of the anaerobic digester(s)
- Anaerobic digester operating temperature
- Method of oxygenation
- Method of external heating of aerobic reactor (if required)
- Feed sludge characteristics
- Ambient temperature

8.2.2 System Retention Times

The user is first asked to select the desired objectives for operation of the dual digester system. In this respect two approaches are available, (1) evaluation of system performance for given aerobic reactor and anaerobic digester retention times (first two choices below) or, (2) calculation of retention time of either the reactor or the digester or both for a defined system performance i.e. sludge quality (3^{rd} to 6^{th} choices below) viz:

- User to select system retention times
- 1 day reactor retention time and 10 day digester retention time
- 1 day reactor retention time and minimum digester retention time
- Minimum reactor retention time and 10 day digester retention time
- Minimum reactor retention time and minimum digester retention time
- Conventional Anaerobic Digestion (aerobic reactor retention time zero)

The **minimum reactor retention time** is defined as the retention time at which the reactor must operate above to ensure that is does not become substrate limited. The method by which the programme calculates this is described in Section 8.3.7 below.

The **minimum digester retention time** is defined as the retention time at which the stability of the final sludge will be equivalent to that of primary sludge (40kg(VS)/m^3) which has undergone conventional anaerobic digestion at 35°C (2.5kg BVS/m^3 remaining). The method by which this is calculated is described in Section 8.3.8 below. The methods by which the volatile solids destruction in both the aerobic reactor and anaerobic digester are determined is described in Sections 8.3.4 and 8.3.5 respectively.

The option of selecting an aerobic reactor retention time of 1 day is provided as this is generally regarded as the optimum retention time for the aerobic stage. It is considered that reducing the reactor retention time below 1 day will increase the risk of washing out the population of thermophilic aerobic bacteria. Likewise the option of an anaerobic digester retention time of 10 days is provided. Below 10 days there is a risk of washing out the methane forming bacteria (sour digesters).

8.2.3 Size, Number, and Shape of the Anaerobic Digester(s)

The user is prompted to enter the capacity (volume) of the anaerobic digester to be constructed (in the case of newly constructed dual digestion plants) or of an existing digester to be upgraded to a dual digestion system. The total number of streams to be simulated is then requested. This factor is included as there may be a number of digesters contributing biogas towards the running of a gas engine (as in the case of Athlone).

The programme then requests information regarding the shape of the digester(s). The following choices are available.

- Anglo-Saxon Contour
- Classic Continental-European Contour
- Egg Shaped

With the digester volume and retention time known, the sludge flow rate through the system is fixed. From the knowledge of aerobic reactor retention time, the size of aerobic reactor(s) can be determined.

Information regarding the shape of the digester is necessary to allow the wall heat losses from the digester to be assessed. The appropriate calculations and a description of the geometry of each of the above shapes is described in Section 8.3.1 below.

8.2.4 Anaerobic Digester Temperature Operating Range

In setting up the model, it was considered that the most appropriate method of operating the dual digestion system is to have the anaerobic digester temperature fixed at either the most appropriate mesophilic or thermophilic temperature for optimum anaerobic biological activity. This is a recommended constraint which the user should place on the model. The input choices are therefore as follows:

- mesophilic (35°C)
- thermophilic (53°C)
- not fixed

Although it is recommended that the digester temperature be fixed at either 35°C or 53°C, the user has the option not to fix the digester temperature (i.e. to allow it be dependent upon a selected aerobic reactor temperature). However, this choice is not recommended due to the risk of the anaerobic digester temperature falling in between the mesophilic and thermophilic temperature regions. This option is not available for the simulation of conventional anaerobic digestion.

8.2.5 Method of Oxygenation

The model presumes that pure oxygen injection will take place. The user is prompted to respond whether air will be used in addition, and if so input the anticipated air flow rate and transfer efficiency of the aeration device. If sufficient heat is provided to the aerobic reactor by the biological heat generated from the air stream and from external heating, then the simulation will calculate that the pure oxygen requirements are zero.

8.2.6 External Heating of the Aerobic Reactor

Four sources of external heating (in addition to internal biological and mechanical heating) of the aerobic reactor are recognised. These are as follows:

- **Interstage Heat Exchange**: The installation of a heat exchanger, between the aerobic reactor and anaerobic digester, to transfer heat from the hot aerobic reactor effluent sludge to the cold reactor influent feed sludge.

- **Heat Recovery from Gas Engine:** The installation of heat exchangers to transfer heat from the gas engine exhaust gasses and cooling water to the cold reactor influent feed sludge.

- **Conventional Boilers:** The use of conventional boilers (hot water heaters), which use biogas as fuel, and heat exchangers to pre-heat the cold reactor influent feed sludge.

- **Afterstage Heat Exchange:** The installation of a heat exchanger to transfer heat from the hot digester effluent sludge (an option available if the digester is operated at thermophilic temperatures) to the cold reactor influent feed sludge.

A discussion of the above four systems is presented in Section 8.3.9 below.

8.2.7 Feed Sludge Characteristics

The model requires knowledge of the type of sludge to be treated by the process. The options available to the user are as follows:

- Primary Sludge
- Primary/Humus Sludge
- Primary/Waste Activated Sludge
- Waste Activated Sludge
- User defined

A breakdown of the different volatile solids fractions for each of the sludge types defined in the programme (listed above) is presented in Table 8.1 below. If the user wishes to define his own VS characteristics then the programme will require information with respect to the percentage volatile solids (VS), and the fraction of biodegradable volatile solids (BVS) and (soluble) readily biodegradable volatile solids (RBVS).

Table 8.1 Volatile Solids Characteristics of the Different Sludge Types Selected by the Simulation Programme to Model the Dual Digestion Process.

Sludge Type	Volatile Solids Fraction of the Total Solids (%)	Biodegradable Fraction of the Volatile Solids (%)	Readily Biodegradable Fraction of the Biodegradable Volatile Solids (%)
Primary Sludge	81	60	20
Primary/Humus Sludge	78	50	10
Primary/ Waste Activated Sludge	75	45	10
Waste Activated Sludge	70	40	0.5

After defining the volatile solids characteristics of the sludge, the user is then asked to give the total solids concentration ($kg(TS)/m^3$) of the sludge to be treated.

8.2.8 Ambient Temperature

The input section of programme concludes with the request for the ambient temperature (°C). This temperature is assigned to the feed sludge, the feed gas (air or oxygen) and the surrounding air (ambient) temperature (for the wall heat loss). Because the feed sludge temperature plays a far more significant role in the heat loss terms than the feed gas and ambient air temperature, the input temperature should be based on the feed sludge temperature. It is recommended when using the programme that the simulation be run using the maximum, minimum and average anticipated ambient (feed sludge) temperatures.

8.3 COMPONENTS OF THE MODEL

This Section describes how the different components of the model which describe the behaviour of the principal system parameters, are calculated. Discussion is included on the logic which is applied when certain parameters are linked, e.g. the interdependency of reactor and digester temperatures and the reduction in biogas production in the digester as oxygen utilisation in the reactor is increased.

8.3.1 Sizing the Aerobic Reactor and Anaerobic Digester

At the start of the programme, the user was asked to either specify the system retention times for the aerobic reactor and the anaerobic digester, or to allow the minimum reactor and/or digester retention times to be calculated based on sludge quality (product) criteria (described in Section 8.2.2 above). To calculate the minimum reactor and/or digester retention times, the programme runs an iterative procedure (for both reactor and digester if required) starting with a reactor retention time of 1.0 day and digester retention time of 10 days until the sludge quality criteria are met (the two iterative processes are described in Sections 8.3.7 and 8.3.8 respectively).

From the system volume V_{dig} (m^3) of the anaerobic digester to be upgraded/constructed at the start of the programme and the selected shape of digester (Anglo-Saxon, European or Egg-shaped), the programme computes, from accepted geometry, the wall surface area of the hot sludge in the digester A_{dig} viz:

Anglo-Saxon (h=r/3) $$A_{dig} = \frac{8\pi}{3}.\left(\frac{3V_{dig}}{\pi}\right)^{2/3} \quad ...m^2 \quad (8.1)$$

Classic Continental European Contour (h=3r) $$A_{dig} = 8\pi.\left(\frac{V_{dig}}{3\pi}\right)^{2/3} \quad ...m^2 \quad (8.2)$$

Egg-Shaped (h=3r) $$A_{dig} = 5\pi.\left(\frac{V_{dig}}{2\pi}\right)^{2/3} \quad ...m^2 \quad (8.3)$$

A value for the wall surface area of the hot sludge in the digester A_{dig} is required to solve the anaerobic digester steady state heat balance, which allows the system temperatures to be determined (see Section 8.3.2).

With the digester capacity and retention time known, the sludge flow rate through the system is calculated. Once the aerobic reactor retention time is known (either selected or obtained by iteration) the size of the aerobic reactor can be determined.

8.3.2 System Temperatures

During the input part of the programme, the user was asked to fix either the digester temperature $T(SL)_d$ (at 35°C or 53°C) or alternately, if the digester temperature is allowed to vary, to fix the aerobic reactor temperature $T(SL)_r$.

If the digester temperature $T(SL)_d$ is fixed, then the digester feed sludge temperature $T(SL)_r$ (normally from the aerobic reactor unless an interstage heat exchanger is installed) is determined by application of the digester steady state heat balance. This heat balance, rearranged in terms of the required digester feed sludge temperature $T(SL)_r$ to maintain a digester temperature of $T(SL)_d$, is as follows:

$$T(SL)_r = \frac{24.R_h.U_0^{dig}.A_{dig}.(T(SL)_d - T_{amb})}{C_p.V_{dig}} + T(SL)_d \quad ...°C \quad (8.4)$$

where:

C_p = The specific heat capacity of the sludge (≈ 4.0 MJ/m^3.°C refer Section 3.7.2)
V_{dig} = The process volume of the digester (m^3)
R_h = The hydraulic retention time in the digester (d)
$T(SL)_r$ = Temperature of the sludge leaving aerobic reactor/entering digester (°C)
$T(SL)_d$ = Temperature of the sludge leaving the anaerobic digester (°C)
U_0^{dig} = Overall heat transfer coefficient (≈ 0.008 MJ/m^2.h.°C)
A_{dig} = Wall area covered by the hot sludge in the digester (m^2)
T_{amb} = Ambient Temperature (°C)

This equation (Eq 8.4) was empirically determined from a number of full scale anaerobic digesters (see Appendix 12).

If the aerobic reactor temperature $T(SL)_r$ is fixed, then the anaerobic digester temperature $T(SL)_d$ is determined by rearranging the digester heat balance in terms of $T(SL)_d$ viz:

$$T(SL)_d = \frac{(C_p.T(SL)_r.V_{dig}) + (24 \times U_0^{dig}.A_{dig}.T_{amb}.R_h)}{(24 \times U_0^{dig}.A_{dig}.R_h) + (C_p.V_{reac})} \quad ...°C \quad (8.5)$$

If interstage cooling is selected, the aerobic reactor temperature $T(SL)_r$ is fixed at 60°C (in order to provide pasteurisation). The term $T(SL)_r$ in Eq 8.4 is then replaced by the term for the temperature of the sludge leaving the heat exchanger (refer Section 8.3.9).

Under South African conditions (with the relatively high ambient temperatures encountered during the summer months +25°C), the use of an interstage heat exchanger will be necessary, at short reactor (<2d) and digester retention times (<15d), if the digester temperature is to be restricted to 35°C (see Appendix 12). Conversely, if no heat exchanger is installed, restricting the digester temperature will mean that the reactor will have to be operated at temperatures below that required for pasteurisation (60°C).

8.3.3 Aerobic Reactor Steady State Heat Balance

Once the aerobic reactor retention time (Section 8.3.1) and the operating temperature (Section 8.3.2) have been defined, the aerobic reactor steady state balance (Eq 3.8) can be solved. The principal purpose in solving the heat balance is to determine the required biological heating rate H_b necessary to maintain the desired reactor temperature $T(SL)_r$. The steady state heat balance is:

$$H_b + H_m = H_s + H_v + H_g + H_w \quad \text{...MJ/h} \quad (3.8)$$

where:

H_b = Rate of biological heat generation
H_m = Rate of heat energy input from the mixing device
H_s = Rate of sensible heat loss with the sludge leaving the reactor
H_v = Rate of vapour heat loss with the effluent gas stream
H_g = Rate of sensible heat loss with the effluent gas stream
H_w = Rate of heat loss through the walls of the reactor

In Chapter 3, expressions were developed to allow each of the heat terms to be determined for operating conditions specific to Athlone. In order to make the model general for application to any sized dual digestion plant, some of the terms in the steady state heat balance derived in Chapter 3 have been modified. The equations used in the simulation model for calculating each of the heat loss and gain terms in the steady state balance are as described below. Where applicable, the modifications made to the equations derived in Chapter 3 are discussed.

Heat Loss Terms

The sludge sensible heat loss rate H_s, gas sensible heat loss rate H_g and the gas vapour heat loss rate H_v terms are not site specific and can be applied unchanged for any size application. These equations, derived in Chapter 3 for oxygenation with air and pure oxygen, are as follows:

The Water Vapour Heat Loss Rate in the Effluent Gas (MJ/h) (Eq 3.133)

$$H_v = 2.38\left(1.21Q(AIR)_{in} + M(O_2)_{in}\right)\cdot\left(\frac{0.62a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{0.62a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}{1240-a\log_{10}\left(9.12-\frac{2307}{273+T(AIR)_{in}}\right)}\right)$$

The Effluent Gas Sensible Heat Loss Rate (MJ/h) (Eq 3.153)

$$H_g = \left(1.21Q(AIR)_{in} + M(O_2)_{in}\right)\cdot\left[0.001\left(T(AIR)_{out} - T(AIR)_{in}\right) + \frac{0.62a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}0.00187T(AIR)_{out}\right]$$

The Sludge Sensible Heat Loss Rate (MJ/h) (Eq 3.163)

$$H_s = 4.04\left[\frac{T(SL)_r Q(SL)_{in}}{24} - \frac{T(SL)_r\left(1.21Q(AIR)_{in} + M(O_2)_{in}\right)}{1000}\cdot\frac{0.62a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)}{760-a\log_{10}\left(8.90-\frac{2239}{273+T(AIR)_{out}}\right)} - \frac{Q(SL)_{in}T(SL)_{in}}{24}\right]$$

where:

$Q(AIR)_{in}$ = The volumetric influent flow rate of **air** to the reactor (m^3(STP)/h)
$M(O_2)_{in}$ = The mass influent flow rate of **pure oxygen** to the reactor (kg(O_2)/h)
$T(SL)_r$ = The aerobic reactor sludge temperature (°C)
$T(SL)_{in}$ = The influent feed sludge temperature (approximated to equal T_{amb}) (°C)
$T(AIR)_{out}$ = The effluent gas temperature (approximated to equal $T(SL)_r$-3) (°C)
$T(AIR)_{in}$ = The influent gas temperature (approximated to equal T_{amb}) (°C)

The wall heat loss rate H_w is site specific and is dependent on the wall surface area covered by the hot sludge in the aerobic reactor. The equation employed for the wall heat loss, is that derived by Messenger *et al* (1992) and equates the wall heat loss rate with reactor area in terms of an overall heat transfer coefficient U_o (see Eq 8.5 below). The reactor wall area is determined, from the geometry of the reactor once the size is known V_{reac} (see Section 8.3.1 above). For simplicity, the model presumes that the reactor is cylindrical in shape, with a height of three times the radius (h=3r; identical to Classic Continental-European Contour). Consequently the wall area A_{reac} is given by:

$$A_{reac} = 8\pi.\left(\frac{V_{reac}}{3\pi}\right)^{2/3} \quad \text{...m}^2 \quad (8.2)$$

The wall heat loss rate H_w (as given by Messenger *et al*) is as follows:

$$H_w = U_0^{reac}.A_{reac}.\left(T(SL)_{reac} - T_{amb}\right) \quad \text{...MJ/h} \quad (8.6)$$

where:

U_0^{reac} = Overall reactor wall heat transfer coefficient (MJ/m^2.°C)

A_{reac} = Wall surface area covered by the hot sludge in the aerobic reactor (m^2)

$T(SL)_{reac}$ = Temperature of the sludge inside the aerobic reactor (°C)

T_{amb} = Ambient temperature i.e. temperature outside the reactor (°C)

In terms of reactor volume V_{reac} the wall heat loss rate is given by:

$$H_w = U_0^{reac} . 8\pi . \left(\frac{V_{reac}}{3\pi} \right)^{2/3} . \left(T(SL)_{reac} - T_{amb} \right) \quad \text{...MJ/h} \quad (8.7)$$

The value for the overall wall heat transfer coefficient U_0^{reac} measured by Messenger *et al* (1992) for the aerobic reactor at Milnerton was U_0^{reac} = 0.00733 MJ/m^2.h. This value has been accepted for use in the model. Consequently Eq 8.7 becomes:

$$H_w = 0.00733 . 8\pi . \left(\frac{V_{reac}}{3\pi} \right)^{2/3} . \left(T(SL)_{reac} - T_{amb} \right) \quad \text{...MJ/h} \quad (8.8)$$

The mechanical heating rate H_m for the Athlone aerobic reactor with pure oxygen oxygenation was determined from the current drawn by the recirculation pumps (see Section 3.9) viz:

$$H_m = 1.65 \times I_{pump} \quad \text{...MJ/h} \quad (3.200)$$

This same relationship was obtained at the Milnerton plant (Messenger *et al*, 1992). In order to make this equation general for any size reactor, a simplistic approximation is made that the current drawn by the recirculation pumps I_{pump} (amps) is equal to the reactor volume V_{reac} (m^3). This approximation is based on the following observation:

At Athlone (phase II)	V_{reac}= 184 m^3	$I_{pump}\approx$ 205 amps	Section 7.3.6; Table 7.11
At Milnerton	V_{reac}= 45 m^3	$I_{pump}\approx$ 44 amps	Messenger *et al* (1992)

Employing the approximation (I_{pump}=V_{reac}), the mechanical heating rate H_m is given by:

$$H_m = 1.65 \times V_{reac} \quad \text{...MJ/h} \quad (8.9)$$

The amount of mixing energy provided by Eq 8.9 is considered sufficient to obtain a pure oxygen transfer efficiency of >80% (achieved at both Athlone and Milnerton) providing the mixing pattern inside the reactor is optimised; this is also required for efficient oxygen transfer (Messenger *et al*, 1992).

The Biological Heating Rate H_b is determined by application of Eq 3.201 where:

$$H_b = Y_h . V_{reac} . (OUR^{AIR} + OUR^{O2}) \quad \text{...MJ/h} \quad (3.201)$$

The oxygen utilisation rate OUR^{AIR} attributable to the utilisation of oxygen from the air stream is calculated from application of the following equation:

$$OUR^{AIR} = OSR^{AIR} . \frac{OTE^{AIR}}{100} = \frac{\rho(AIR) . f_{O2}^{AIR} . Q(AIR)_{in} . OTE^{AIR}}{V_{reac} . 100} \quad \text{...kg}(O_2)/m^3.h \quad (8.10)$$

where:

OSR^{AIR} = The oxygen supply rate of oxygen from the air stream (kg(O_2)/m^3.h)
OTE^{AIR} = The oxygen transfer efficiency of the air oxygenation device (%)
$\rho(AIR)_{in}$ = The density of atmospheric air at STP (=1.205 kg/m^3)
f_{O2}^{AIR} = The mass fraction of oxygen in atmospheric air (=0.23 kg/kg)
$Q(AIR)_{in}$ = The volumetric influent flow rate of air to the reactor (m^3(STP)/h)
V_{reac} = The process volume of the aerobic reactor (m^3)

Both the influent air flow rate $Q(AIR)_{in}$ and the transfer efficiency of the air oxygenation device OTE^{AIR} were given by the user at the start of the simulation. The reactor process volume V_{reac} is computed by the programme (see Section 8.3.1).

The oxygen utilisation rate OUR^{O2} attributable to the utilisation of oxygen from the pure oxygen injected is calculated from application of the following equation:

$$OUR^{O2} = OSR^{O2} . \frac{OTE^{O2}}{100} = \frac{M(O_2)_{in} . OTE^{O2}}{V_{reac} . 100} \quad \text{...kg}(O_2)/m^3.h \quad (8.11)$$

where:

OSR^{O2} = The oxygen supply rate of the pure oxygen stream (kg(O_2)/m^3.h)
OTE^{O2} = The oxygen transfer efficiency of the pure oxygen injection device (accepted to be = 80%. Sufficient mixing energy is provided to ensure this; refer previous sub-section on mechanical heat input)
$M(O_2)_{in}$ = The mass flow rate of pure oxygen entering the reactor (kg(O_2)/h)

Solving the Aerobic Reactor Steady State Heat Balance

The aerobic reactor steady state heat balance (Eq 3.8) is solved by an iterative process (see Figure 8.1). The programme increases the flow rate of pure oxygen to the reactor $M(O_2)_{in}$ from 0.0 kg(O_2)/h in small increments (0.1 kg(O_2)/h) until the heat input to the reactor (H_b+H_m) is equal to the heat losses (H_s+H_v+H_g+H_w). The computation is then terminated. At this point, the required pure oxygen injection rate $M(O_2)_{in}$ and the biological heating rate H_b is known.

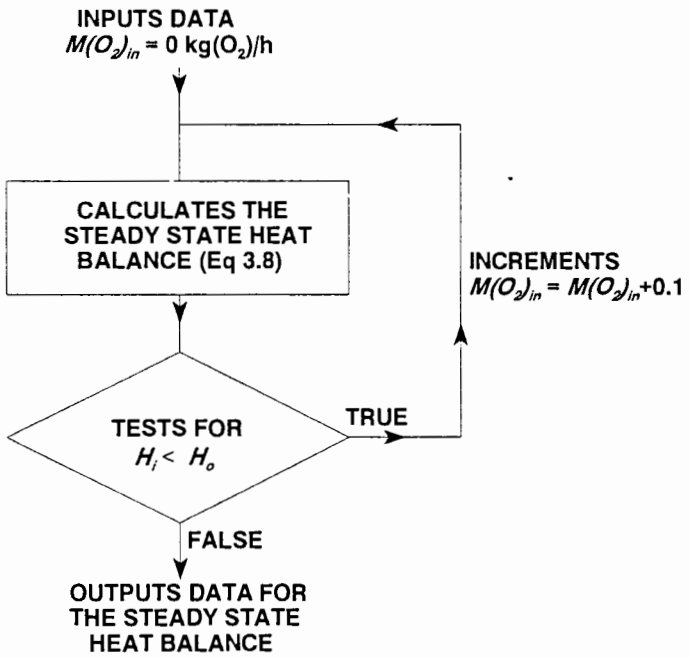


Figure 8.1 The Iterative Process Applied in the General Dual Digestion Model to Calculate the Aerobic Reactor Steady State Heat Balance.

8.3.4 Volatile Solids Destruction in the Aerobic Reactor

From the results of the investigation at Athlone (both phase I and phase II), convincing evidence was found of a link between the rate of oxygen utilisation $M(O_2)_{ut}^{AIR+O2}$ ($=V_{reac}\times OUR^{AIR+O2}$) and the rate of volatile solids destruction $M(VS)_{dest}^{reac}$ in the aerobic reactor (see Section 7.5). The ratio $M(O_2)_{ut}^{AIR+O2}$ to $M(VS)_{dest}^{reac}$ was found to be close to the COD/VS ratio (f_{ovs}) of the influent feed sludge (which for Athlone f_{ovs}= 1.70 kg(O_2)/kg(VS); see Table 4.1). Consequently the rate of volatile solids destruction in the aerobic reactor is approximated in the programme by the following equation:

$$M(VS)_{dest}^{reac} = \frac{V_{reac}.OUR^{AIR+O2}}{f_{ovs}} \quad \text{...kg(VS)/h} \quad (8.12)$$

8.3.5 Volatile Solids Destruction in the Anaerobic Digester

The rate of volatile solids destruction inside the digester $M(VS)_{dest}^{dig}$ is calculated with the aid of the digester model in Appendix 11. This model, which simulates the breakdown of volatile solids in the anaerobic digestion process, accepts the six distinct anaerobic conversion processes depicted in Figure 11a.1 as described by Guger and Zehnder (1983). However, for simplicity the conversion of acetate to methane (acetogenic methanogenesis) is considered to be the rate limiting step in the process. The rate of acetogenic methanogenesis is modelled on the kinetic equation developed by Grau *et al*

(1975) (refer Eq11a.1; Section 11a.3) and the stoichiometry follows that of Siegrist *et al* (1993) to describe the breakdown of the influent COD (primary sludge VS) through the different metabolic pathways in the process prior to acetogenic methanogenesis (refer Table 11a.1; Section 11a.3). Based on Eq 11a.20, the rate of volatile solids destruction in the digester $M(VS)_{dest}^{dig}$ in units of kg(VS)/h is given by:

$$M(VS)_{dest}^{dig} = \frac{V_{dig}.f_{ovs}.[BVS]_{in}}{24.R_h}\left(0.823 - 0.582\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) \quad \text{...kg(VS)/h} \qquad (8.13)$$

where:

$[BVS]_{in}$ = Concentration of Biodegradable Volatile Solids (BVS) in the feed sludge to the digester (ex. aerobic reactor) (kg(VS)/m^3)

$\hat{\mu}$ = Maximum specific growth rate of the methanogenic organisms (d^{-1})

b_T = Decay rate of the methanogenic organisms (d^{-1})

R_h = Solids retention time in the digester (d)

V_{dig} = Anaerobic digester process volume (m^3)

f_{ovs} = COD/VS ratio of the sludge (kg(O)/kg(VS))

The anaerobic digestion model developed in Appendix 11 is also capable of predicting biogas production rates and the concentrations of methane and carbon dioxide in the biogas (see Table 11a.5; Section 11a.4). The relevant formulae developed in this Section (11a.4) are employed in the general dual digestion model to predict biogas production rate and composition.

8.3.6 Final Sludge Stability

Volatile solids are destroyed in both the aerobic reactor and anaerobic digester. The rates at which this destruction takes place are described in Sections 8.3.4 and 8.3.5 above respectively. The simulation model describes the stability of the final sludge from the dual digestion system in terms of the following parameters:

- The percentage volatile solids destruction achieved in the system. Heidman (1989) considers that >38% volatile solids destruction is necessary for satisfactory sludge stabilisation.

- The concentration of biodegradable volatile solids (BVS) remaining in the sludge after treatment. Comparison is made with the value 2.5kg(BVS)/m^3. This is the level of BVS remaining in primary sludge (initial concentration 40kg(TS)/m^3) after conventional mesophilic (35°C) anaerobic digestion at a retention time of 20 days (as predicted by the anaerobic digestion model described in Appendix 11).

- The predicted peak specific oxygen utilisation rate ***SOUR*** ($g(O_2)/kg(TSS).h$) which occurs during the first 24 hours of the ***SOUR*** test. This prediction is based on an application of aerobic activated sludge kinetics as described by Dold *et al* (1991), details of which are given in Appendix 11. Details of the actual ***SOUR*** test are provided in Appendix 7.

8.3.7 Minimum Aerobic Reactor Retention Time

The user has the option to request the programme to predict the aerobic reactor retention time above which it is necessary to operate to ensure that the reactor does not become substrate limited, i.e. that there is sufficient readily biodegradable volatile solids (RBVS) available to match the required oxygen utilisation rate (in terms of Eq 8.12 above).

Under substrate limiting conditions, the rate of volatile solids destruction $M(VS)_{dest}$ under thermophilic aerobic conditions (at relatively long retention times, >3days) can be predicted using the kinetics of Andrews and Kambhu (1971) (refer Section 3.11.4). For application at shorter retention times (<3 days) the kinetic equation (Eq 3.212) has been modified. At the shorter retention times, it is considered that the rate limiting step in the destruction of VS is that of enzyme hydrolysis i.e. the conversion of particulate biodegradable volatile solids (BVS) to soluble readily biodegradable volatile solids (RBVS). Applying a first order rate equation to describe the rate of hydrolysis, and making allowance for the RBVS entering the reactor in the feed sludge, the rate of VS destruction is given by:

$$M(VS)_{dest} = \frac{k_h.[BVS]_{in}.V_{reac}}{24(1 + R_h.k_h)} + \frac{V_{reac}.[RBVS]_{in}}{24.R_h} \quad \text{...kg(VS)/h} \quad (8.14)$$

where:

$M(VS)_{dest}$ = The rate of volatile solids destruction (kgVS/h)
k_h = The hydrolysis (volatile solid destruction) rate coefficient (/d)
R_h = The hydraulic retention time (d)
V_{reac} = The effective process volume of the aerobic reactor (m^3)
$[BVS]_{in}$ = The biodegradable VS conc. in the influent sludge ($kgBVS/m^3$)
$[RBVS]_{in}$ = The readily biodegradable VS conc. in the influent sludge ($kgRBVS/m^3$)

In the model, the hydrolysis rate coefficient k_h is assigned the same value as that taken for the destruction rate coefficient k_d in the kinetics of Andrews and Kambhu (1971). This value, which is temperature dependent, is determined from an application of the following equation (linear fit to the graphics presented by Fuggle and Spensley, 1985):

$$k_h = 0.0025 \times T(SL)_r + 0.125 \quad \text{(valid in the region 40°C to 60°C)} \quad .../d \quad (8.15)$$

At the point at which the reactor becomes substrate limited, Equations Eq 8.12 (for oxygen limiting conditions) and Eq 8.14 (for substrate limiting conditions) will both be valid for describing the rate of volatile solids destruction $M(VS)_{dest}$, and therefore be equal to each other:

Oxygen Limiting = Substrate Limiting

$$M(VS)_{dest} = \frac{V_{reac}.OUR^{AIR+O2}}{f_{ovs}} = \frac{k_h.[BVS]_{in}.V_{reac}}{24(1 + R_h.k_h)} + \frac{V_{reac}.[RBVS]_{in}}{24.R_h} \quad ...kg(VS)/h \quad (8.16)$$

At this transition point, OUR^{AIR+O2} is at its maximum rate OUR^{max} (when the aerobic reactor operates under substrate limiting conditions $OUR < OUR^{max}$). A value for OUR^{max} can therefore be determined by rearranging Eq 8.16 accordingly:

$$OUR^{max} = \frac{f_{ovs}.k_h.[BVS]_{in}}{24(1 + R_h.k_h)} + \frac{f_{ovs}.[RBVS]_{in}}{24.R_h} \quad ...kg(O_2)/m^3.h \quad (8.17)$$

By solving the aerobic reactor steady state heat balance (refer Section 8.3.3 above), the programme is able to compute the required OUR (i.e. biological heating rate H_b) for a specific reactor temperature $T(SL)$, and defined retention time R_h. In predicting the retention time at which substrate limitation occurs, the programme starts an iterative process (refer Figure 8.2 below). The initial retention time is taken as 1 day. If the calculated $OUR > OUR^{max}$ then the retention time is increased (+0.01 day). The iteration continues until $OUR=OUR^{max}$, at which point the programme gives the steady state data.

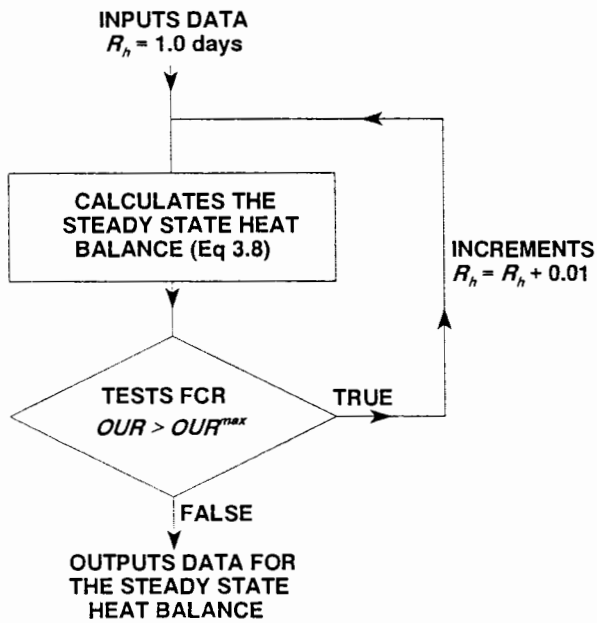


Figure 8.2 **The Iterative Process Applied in the General Dual Digestion Model to Calculate the Minimum Aerobic Reactor Retention Time (The Point at which Substrate Limitation Occurs).**

8.3.8 Minimum Anaerobic Digester Retention Time

The minimum anaerobic digester retention time is defined as the retention time at which the stability of the final sludge will be equivalent to that of primary sludge (40kg(VS)/m^3) which has undergone conventional (20d) anaerobic digestion at 35°C (2.5kg BVS/m^3 remaining). The method of predicting the volatile solids destruction in the digester has been described in Section 8.3.5 above (presented in detail in Appendix 11). The simulation model predicts the minimum anaerobic digester retention time by starting an iterative process with the digester retention time fixed at 10 days (see Figure 8.3). The volatile solids destruction in the digester and the concentration of biodegradable volatile solids $[BVS]_{out}$ in the effluent sludge are then determined. If the calculated concentration of $[BVS]_{out} > 2.5$ kg(BVS)/m^3 then the retention time is increased by 0.1 days. The iteration then continues until the criterion is met.

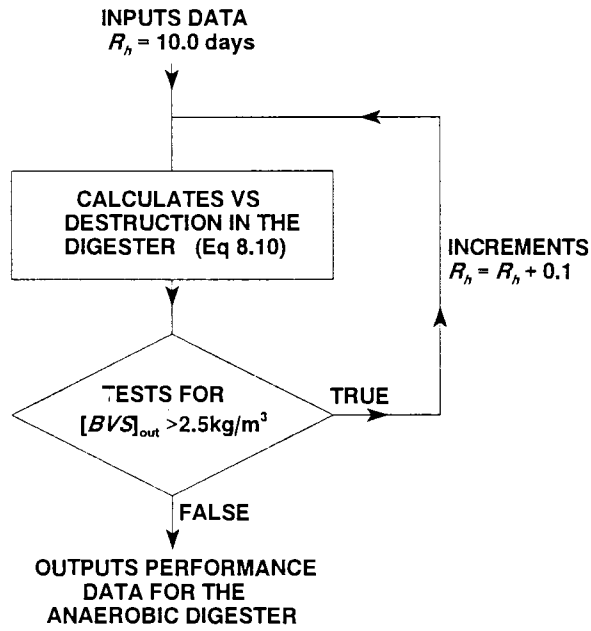


Figure 8.3 The Iterative Process Applied in the General Dual Digestion Model to Calculate the Minimum Anaerobic Digester Retention Time to Produce a Stable Sludge

8.3.9 External Heating

The rate of pure oxygen injection $M(O_2)_{in}$, required to maintain a specific reactor temperature, can be reduced if an additional source of heating is supplied. The two obvious sources of external heat are:

- The recovery of sensible heat from either the hot aerobic sludge or the hot digester sludge (if operated in the thermophilic mode).

- The generation of heat from the burning of the biogas generated during anaerobic digestion. Typically either through the use of conventional boilers (hot water systems) or the recovery of heat from an installed gas engine.

The simulation model recognises 4 options (encompassing the above), which are as follows:

Option 1: **Interstage Heat Exchange**: The installation of a heat exchanger between the aerobic reactor and anaerobic digester to transfer heat from the hot aerobic reactor effluent sludge to the cold reactor influent feed sludge H_{x1}^{trans}.

Option 2: Heat Recovery from a Gas Engine: The installation of heat exchangers to transfer heat from the exhaust gasses and cooling water of an installed gas engine to the cold reactor influent feed sludge H_{x2}^{trans}.

Option 3: Conventional Boilers: The use of conventional boilers (hot water heaters), which use biogas as fuel, and heat exchangers to pre-heat the cold reactor influent feed sludge H_{x3}^{trans}.

Option 4: **Afterstage Heat Exchange:** The installation of a heat exchanger to transfer heat from the hot anaerobic digester effluent sludge (an option available if the digester is operated at thermophilic temperatures) to the cold reactor influent feed sludge H_{x4}^{trans}.

The user is asked to make the required choice during the input part of the programme (see Section 8.2.6). The method of computation for each option is described below.

Option 1: Interstage Heat Exchange

A direct sludge|sludge heat exchanger is installed between the aerobic reactor and anaerobic digester to transfer sensible heat from the hot aerobic sludge to the cold influent feed sludge (see Figure 8.4 below).

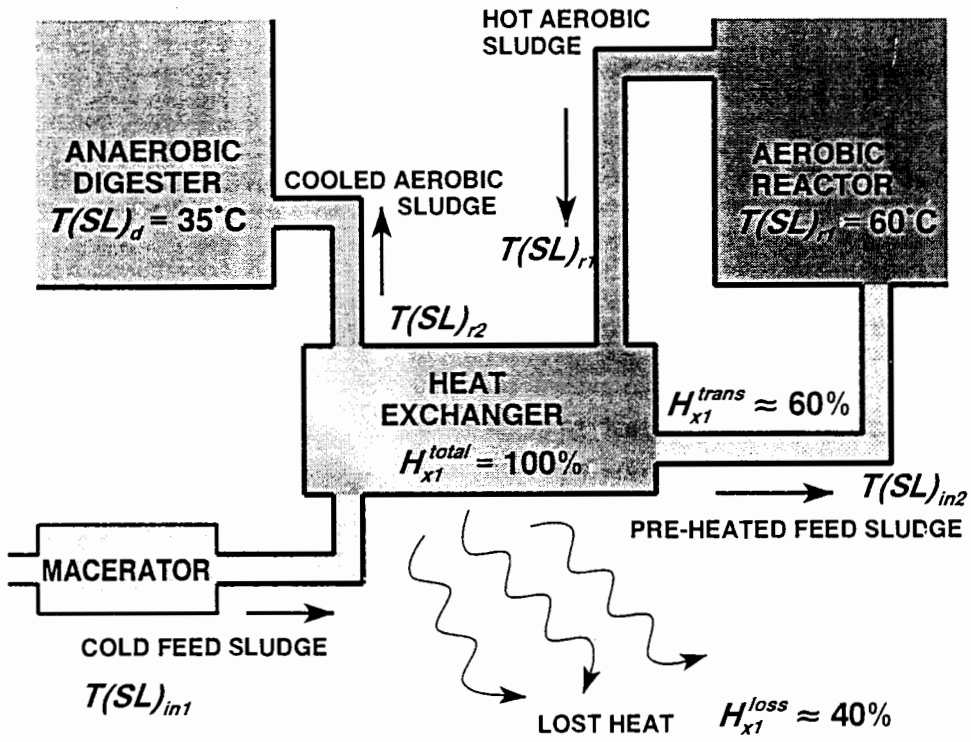


Figure 8.4 Schematic of a Dual Digestion Plant with Interstage Heat Exchange

For this option the model assumes that the reactor temperature is controlled at $T(SL)_{r1}$ = 60°C (to achieve pasteurisation) with the digester temperature at $T(SL)_d$ = 35°C (for efficient mesophilic stabilisation). The required temperature of the sludge entering the digester $T(SL)_{r2}$ is determined from an application of the digester steady state heat balance (refer Section 8.3.2; Eq 8.4). The rate at which sensible heat needs to be transferred from the hot aerobic sludge to the cold feed sludge (including heat losses) is given by:

$$H_{x1}^{total} = C_p \cdot \frac{Q(SL)_{in}}{24} \cdot (T(SL)_{r1} - T(SL)_{r2}) \quad \text{...MJ/h} \quad (8.18)$$

where:

H_{x1}^{total} = The rate at which sensible heat is given up by the hot aerobic sludge through the heat exchanger (MJ/h)

C_p = The specific heat capacity of the feed sludge (=4.0 MJ/m^3.°C)

$Q(SL)_{in}$ = The flow rate of sludge through the process (m^3/d)

$T(SL)_{r1}$ = The temperature of the sludge leaving the aerobic reactor/entering the heat exchanger (=60°C)

$T(SL)_{r2}$ = The temperature of the sludge entering the anaerobic digester/ leaving the heat exchanger (°C)

Typically, sludge | sludge heat exchangers are approximately 60% efficient in the transfer of heat (Bruce and Oliver, 1987). Consequently, in the model, the rate of heat transfer to the cold feed sludge is calculated by:

$$H_{x1}^{trans} = \frac{60}{100}.H_{x1}^{total} = \frac{60}{100}.C_p.\frac{Q(SL)_{in}}{24}.(T(SL)_{r1} - T(SL)_{r2}) \quad \text{...MJ/h} \quad (8.19)$$

where:

H_{x1}^{trans} = The rate of heat transfer to the cold feed sludge (MJ/h)

While interesting to know, it is not necessary for the heat balance to calculate the increase in the feed sludge temperature (which can be done with the aid of Eq 8.21 below). All that is required is the rate of sensible heat transfer to the sludge $H(SL)_{x1}^{trans}$ which Eq 8.15 provides. The principal effect of transferring heat to the incoming sludge is that the required biological heating rate H_b is reduced by $H(SL)_{x1}^{trans}$.

The sensible heat gain by the incoming feed sludge $H(SL)_{x1}^{trans}$ can be expressed in the following form:

$$H_{x1}^{trans} = C_p.\frac{Q(SL)_{in}}{24}.(T(SL)_{in2} - T(SL)_{in1}) \quad \text{...MJ/h} \quad (8.20)$$

where:

$T(SL)_{in2}$ = The temperature of the feed sludge leaving the heat exchanger/ entering the aerobic digester (°C)

$T(SL)_{in1}$ = The temperature of the feed sludge entering the heat exchanger, accepted by the model to be at ambient temperature (= T_{amb}°C)

The temperature of the cold feed sludge leaving the heat exchanger $T(SL)_{in2}$ can be determined by rearranging Eq 8.20 viz:

$$T(SL)_{in2} = \frac{H_{x1}^{trans}.24}{C_p.Q(SL)_{in}} + T(SL)_{in1} \quad \text{...°C} \quad (8.21)$$

It is important to stress that the calculated heat exchanger outlet temperatures [$T(SL)_{in2}$ and $T(SL)_{r2}$] are the required operating temperatures for the heat exchanger in order for the dual digester to operate as prescribed (reactor at 60°C, digester at 35°C). In practice, the actual heat exchanger outlet temperatures will depend on heat exchanger operating conditions, principally:

- Overall heat transfer coefficient
- Area of heat exchanger
- Contact time

Accordingly, the heat exchanger should be designed and operated to achieve the required sludge temperatures. The types of sludge | sludge heat exchangers in common use range from simple tube-in-tube or stirred tank-in-tank to specialised spiral type heat exchangers. The major operational problems associated with the spiral type of exchanger relate to the pumping of the sludge through the narrow passageways, maintaining the required velocities, controlling fouling and maintaining a reasonable overall heat transfer efficiency (Bruce and Oliver, 1987). **The installation of a sludge macerator and efficient screening system upstream of the dual digestion system is strongly recommended** to minimize these problems.

Option 2: The Recovery of Heat from an Installed Gas Engine

The recovery of heat from an installed gas engine is a common method of heating sludge for digestion (Bruce and Oliver, 1987). A schematic of a gas engine with a water | sludge heat exchanger is shown in Figure 8.5 below. Note that it is possible to recover heat from both the exhaust gases and the engine cooling water.

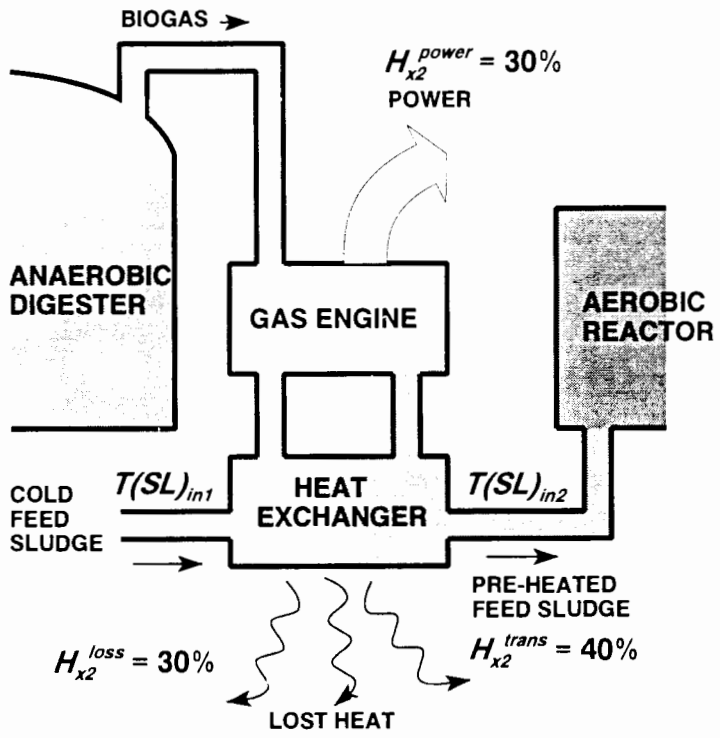


Figure 8.5 Schematic of a Dual Digestion Plant with Gas Engine

The gas engine will require a certain quantity of biogas to enable it to run for 24 hours per day. Accepting a 40% efficiency in the transfer of energy from the combustion of methane to the influent feed sludge in the form of sensible heat (via the heat exchanger), the following set of equations are employed in the model to define the sludge heating rate H_{x2}^{trans}.

The Heat of Combustion of Methane

$$Y_{CH_4} = 0.882 \quad \text{...MJ/mol}(CH_4) \quad (8.22)$$

The Rate of Methane Production

The molar rate of methane production $n(CH_4)_{gen}$ is estimated using the anaerobic digestion model described in Appendix 11 (refer Table 11a.5). In terms of the digester feed sludge biodegradable volatile solids concentration $[BVS]_{in}$ the molar rate of methane production $n(CH_4)_{gen}$ in units of mol(CH_4)/h is given by:

$$n(CH_4)_{gen} = \frac{f_{ovs}.V_{dig}.[BVS]_{in}}{24.R_h}.\left(12.83 - 9.08\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) \quad \text{...mol}(CH_4)/\text{h} \quad (8.23)$$

where:

$[BVS]_{in}$ = Concentration of Biodegradable Volatile Solids (BVS) in the feed sludge to the digester (ex. aerobic reactor) (kg(VS)/m^3)

$\hat{\mu}$ = Maximum specific growth rate of the methanogenic organisms (d^{-1})

b_T = Decay rate of the methanogenic organisms (d^{-1})

R_h = Solids retention time in the digester (d)

V_{dig} = Anaerobic digester process volume (m^3)

f_{ovs} = COD/VS ratio of the sludge (kg(O)/kg(VS))

Energy Specifications of the Gas Engine

An input power rating in kW units (≡ 1/3.6 MJ/h) is usually provided with the installed gas engine. For Athlone this value is 1300kW (or equivalently 4680 MJ/h). In units of MJ/h, the input power requirement of the gas engine H_{x2}^{req} must be equivalent the rate of heat release through the combustion of methane (biogas);

$$H_{x2}^{req} = Y_{CH_4}.n(CH_4)^{req} \quad \text{...MJ/h} \quad (8.24)$$

where:

H_{x2}^{req} = Heating requirement for the gas engine (MJ/h)

$n(CH_4)^{req}$ = Required rate of methane production to run the gas engine (mol(CH_4)/h)

It follows therefore that to run the gas engine for 24 hours:

$$\boldsymbol{n}(CH_4)_{gen} > \boldsymbol{n}(CH_4)^{req} \quad \text{...mol}(CH_4)/h \quad (8.25)$$

If the rate of methane production $\boldsymbol{n}(CH_4)_{gen}$ falls below the required production rate $\boldsymbol{n}(CH_4)^{req}$ then the time period for which it will be possible to run the gas engine will fall below 24hrs per day viz:

$$\boldsymbol{t}_{eng}^{run} = 24.\frac{\boldsymbol{n}(CH_4)_{gen}}{\boldsymbol{n}(CH_4)^{req}} \quad \text{(up to a maximum value of 24)} \quad \text{...h} \quad (8.26)$$

If $\boldsymbol{t}_{eng}^{run} > 24$h then surplus gas is generated and the proportions of the total gas generated that are utilised and wasted (surplus) are given by the model as percentages.

The Transfer of Energy

Accepting the efficiencies for the system quoted in Figure 8.5 above (40% transfer of sensible heat, 30% transfer to mechanical/electrical energy, 30% heat losses; derived from data provided by the specifications of the Athlone gas engine and Bruce and Oliver ,1987), it follows that the average rate of sensible heat transfer to the cold reactor influent feed sludge $\boldsymbol{H}_{x2}^{trans}$ is given by:

$$\boldsymbol{H}_{x2}^{trans} = 0.4 \times \frac{\boldsymbol{t}_{eng}^{run}}{24}.\boldsymbol{Y}_{CH_4}.\boldsymbol{n}(CH_4)^{req} \quad \text{(the maximum value of } \boldsymbol{t}_{eng}^{run} = 24) \quad \text{...MJ/h} \quad (8.27)$$

Likewise, the rate of mechanical/electrical energy generation $\boldsymbol{H}_{x2}^{power}$ is given by:

$$\boldsymbol{H}_{x2}^{power} = 0.3 \times \frac{\boldsymbol{t}_{eng}^{run}}{24}.\boldsymbol{Y}_{CH_4}.\boldsymbol{n}(CH_4)^{req} \quad \text{(the maximum value of } \boldsymbol{t}_{eng}^{run} = 24) \quad \text{...MJ/h} \quad (8.28)$$

During the input stage of the programme, the user has the choice to restrict the quantity of sensible heat to be transferred to the feed sludge (as a percentage of the total available). The value for $\boldsymbol{H}_{x2}^{trans}$ (Eq 8.27) is then reduced accordingly.

Option 3: Conventional Hot Water Boilers:

Conventional hot water boilers have traditionally been the method used for heating conventional anaerobic digesters. Through the combustion of methane (biogas) water is raised to a temperature of around 80°C and circulated in a closed loop through a water | sludge heat exchanger. The efficiency of sensible heat transfer into the sludge is typically 60-65% (Bruce and Oliver, 1987).. As dual digestion is an option for upgrading conventional digestion plants, it is logical to expect that in a number of instances hot water boilers will be available to provide an additional heat source if desired. Figure 8.6

shows a schematic of a dual digestion plant fitted with a hot water boiler and heat exchanger.

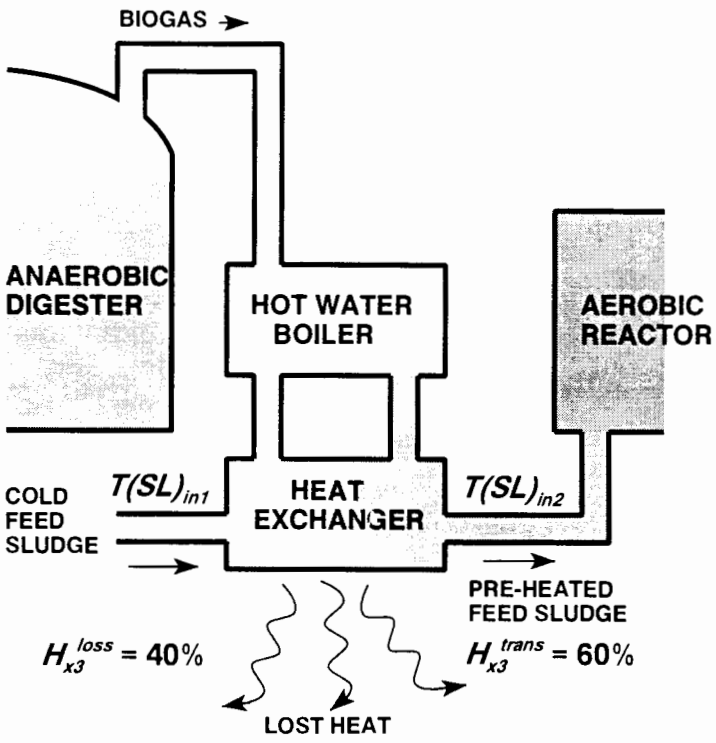


Figure 8.6 Schematic of a Dual Digestion Plant with Hot Water Boiler

The simulation model accepts a figure of 60% for the efficiency of transfer of sensible heat. Consequently, the rate of sensible heat transfer H_{x3}^{trans} to the feed sludge is given by:

$$H_{x3}^{trans} = 0.6 Y_{CH_4}.n(CH_4)_{gen} \quad \text{...MJ/h} \quad (8.29)$$

The user has the choice to limit the amount of sensible heat transferred (as a percentage of the total available). This may be required as a result of (a) limitations to the hot water boiler/heat exchanger system or (b) a certain level of oxygen injection may be desired.

Option 4: Afterstage Heat Exchanger

For this option, a direct sludge|sludge heat exchanger is installed after the anaerobic digester to transfer sensible heat from the hot digester effluent sludge to the cold reactor influent feed sludge (Figures 8.7). The principals involved in the installation of an after-stage heat exchanger are similar to those described above for the inter-stage exchanger (Option 1). However, as there is now no restriction on the outlet hot sludge temperature (ex anaerobic digester) $T(SL)_{d2}$, this temperature and the outlet cold sludge temperature (ex feed sludge) $T(SL)_{in2}$ can be allowed to reach equilibrium. i.e.

$$T(SL)_{in2} = T(SL)_{d2} \quad \text{...°C} \quad (8.30)$$

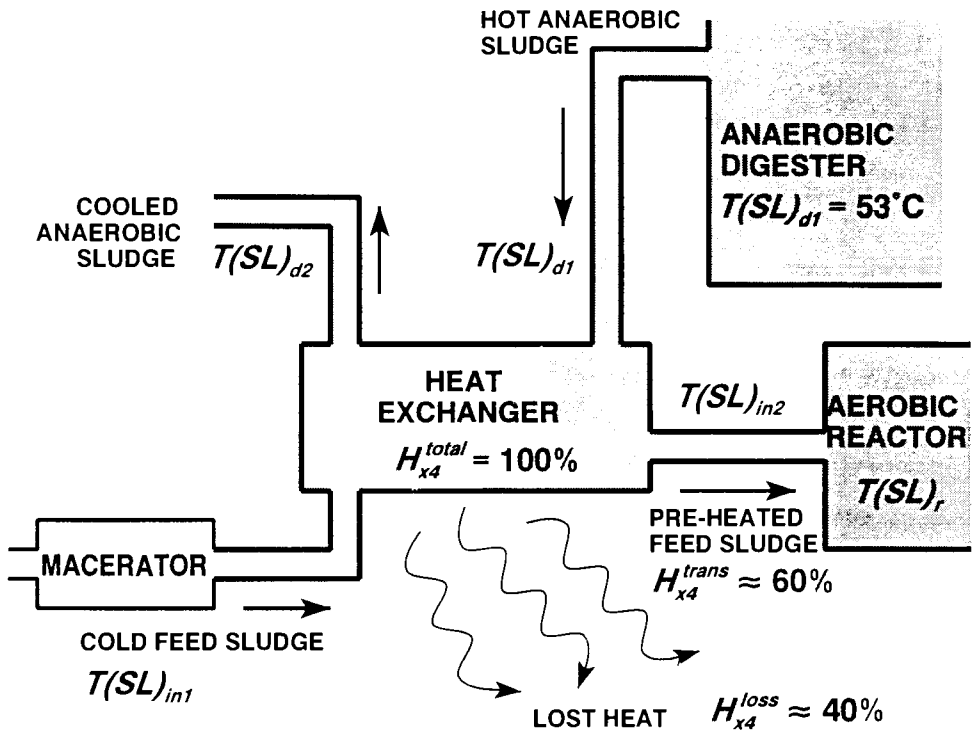


Figure 8.7 Schematic of a Dual Digestion Plant with Afterstage Heat Exchange

The rate of heat transfer from the hot digester effluent sludge to the cold reactor feed sludge is determined by applying a steady state heat balance across the heat exchanger and accepting a 60% efficiency in the transfer of heat (Bruce and Oliver, 1987) viz:

$$H_{x4}^{trans} = \frac{60}{100}.H_{x4}^{total} = 0.6 \times C_p.\frac{Q(SL)_{in}}{24}.(T(SL)_{d1} - T(SL)_{d2}) = C_p.\frac{Q(SL)_{in}}{24}.(T(SL)_{in2} - T(SL)_{in1}) \quad ...MJ/h \quad (8.31)$$

where:

H_{x4}^{total} = The rate at which sensible heat is given up by the hot digester sludge through the heat exchanger (MJ/h)

$T(SL)_{d1}$ = The temperature of the sludge leaving the anaerobic digester/entering the heat exchanger (=53°C)

$T(SL)_{d2}$ = The temperature of the sludge leaving the heat exchanger (=$T(SL)_{in2}$ °C)

The temperature of the two sludge streams leaving the heat exchanger ($T(SL)_{in2}$ and $T(SL)_{d2}$; which are presumed equal- Eq 8.30) is calculated by substituting Eq 8.30 into the heat balance equation (Eq 8.31) rearranged in terms of $T(SL)_{in2}$ (or equivalently $T(SL)_{d2}$):

$$T(SL)_{in2} = T(SL)_{d2} = 0.375\,T(SL)_{d1} + 0.625\,T(SL)_{in1} \quad ...°C \quad (8.32)$$

The rate of heat transfer to the cold reactor feed sludge H_{x4}^{trans} is obtained by substituting Eq 8.32 back in the exchanger heat balance (Eq 8.31) viz:

$$H_{x4}^{trans} = 0.375 \times C_p . \frac{Q(SL)_{in}}{24} . (T(SL)_{dl} - T(SL)_{inl}) \quad \text{...MJ/h} \quad (8.33)$$

8.3.10 Dual Digestion Cost Analysis

In order to make a cost comparison between different dual digester system configurations and conventional anaerobic treatment, the model makes an assessment of costs (both operating and capital) at the end of each simulation. To combine operating and capital costs (to give a total system cost p.a.), the annual repayment on a capital loan (over 15 years at 18% interest) is added to the projected annual operating cost in 7½ years time (at 15% inflation ≡ 2.85× current operating cost); this being the mid-term of the loan period. The cost assessment for each item is based on the following considerations:

Operational Costs

The operational costs are considered to be those incurred through (a) the use of pure oxygen for biological heating of the aerobic reactor, (b) electricity consumption for air production (if required) and mixing of the aerobic reactor, and (c) transportation costs (to a suitable disposal site) if the sludge is not pasteurised. The oxygen and electricity costs are based on the following (prices quoted are for 1995, Cape Town):

Pure Oxygen:	0.45	SA Rands/kg(O_2)
Electricity:	0.20	SA Rands/kWh

The quantity of pure oxygen required is determined from the steady state heat balance. The electricity requirements for air supply and mixing are based on the application of the following formulae:

Air Supply:	$0.025Q(AIR)_{in}$	kW/m^3(AIR)	...Based on actual data for Athlone
Reactor Mixing:	$0.46V_{reac}$	kW/m^3	...Based on criterion $I_{pump}=V_{reac}$ (Section 8.3.3)
Digester Mixing:	$0.05V_{dig}$	kW/m^3	...Typical design value

The total disposal cost (1995) is calculated from a unit cost of transporting one truck containing $10m^3$ of dried sludge (50% total solids; after 14 days on drying beds- not included in costing) a distance of 20km (40km round trip):

Cost per km	4.00	SA Rands/km
Transportation unit cost	0.032	SA Rands/kg(TS)

If a gas engine is operated, operating costs are recovered. This cost saving is reflected as an electrical cost saving:

Costs Recovered:	0.72	SA Rands/MJ ...equivalent to 0.20 SA Rands/kW

Capital Costs:

Capital costs are considered to be incurred for the following items (if incorporated):

- Aerobic reactor (including recirculation pipework)
- Reactor recirculation pumps
- Compressor for the supply of air
- Anaerobic digester (inc. mixing pumps, biogas recirculation network, gas holder)
- Conventional hot water boiler
- Gas engine
- Heat exchanger system

The following formulae are applied to estimate the capital costs for each of the above items. It is recognised that the data obtained from an application of this nature, will in all likelihood vary widely from actual capital costs incurred (which will be strongly dependent on specific site applications). However, the information obtained will allow a relative comparison between different system configurations and an approximate assessment of the cost of the various dual digestion systems to be made.

Aerobic reactor (concrete structure) = $1.45[0.001V_{reac}]^{0.63}$

where: cost is in R000,000's (millions)

V_{reac} = the reactor process volume (m^3)

The function $[0.001V_{reac}]^{0.63}$ is that employed by Wolinski (1984). The factor 1.45 is applied to give a cost of R500,000 for a $184m^3$ reactor (estimated).

Recirculation Pumps (aerobic reactor) = R2000 /kW

Compressor (for the supply of air) = R2500 /kW

Anaerobic digester (concrete structure) = $1.94[0.001 V_{dig}]^{0.63}$
where: cost is in R000,000's (millions)
V_{dig} = the digester process volume (m^3)
The factor 1.94 is applied to the Wolinski (1984) function to give a 1995 cost of R3,000,000 for a $2000m^3$ digester, which is a reasonable current (1995) capital estimated. Note that this figure includes mixing equipment, gas recirculation equipment and gas holder costs.

Conventional hot water boiler = R500,000 per 1000 m^3(BIOGAS)/d
based on the estimated capital cost for a boiler with capacity of 1000 m^3(BIOGAS)/d.
Note: BIOGAS includes methane and carbon dioxide.

Gas Engine = R1,000,000 per 1000 m^3(BIOGAS)/d
based on the estimated capital cost of R5,000,000 for an engine using $5000m^3$(BIOGAS)/d

Heat Exchangers = R200 per kW of heat transferred.
The heat transferred is proportional to the area (size) of heat exchanger required.

8.4 INFORMATION PROVIDED BY THE MODEL

8.4.1 Summary of the Information Provided

On completion of the simulation, the results generated by the programme are printed to the screen. The user is given the option to make a hard copy of the output (Figure 8.8). The summary of the information provided is listed below:

- **System Data:** Sludge volume treated. Reactor and digester process capacities and retention times. Method of oxygenation. Methods of external heating employed. Temperature of feed, reactor and digester sludges.

- **Sludge Quality:** Total Solids, Volatile Solids and Biodegradable Volatile Solids concentrations (kg/m^3) for the influent, reactor and digester sludges. The percentage volatile solids removed at each stage in the system.

- **Oxygen Utilisation Rates:** The ***OUR*** attributable to pure oxygen and air (if selected). The maximum ***OUR*** for the sludge type treated. The required rate of pure oxygen injection ($kg(O_2)/h$).

- **Biogas Data:** The rate of biogas production in the anaerobic digester. The quantities of biogas utilised and wasted in the process. The concentrations of methane and carbon dioxide in the biogas.

- **Energy Balance Aerobic Reactor :** The calculated values for each of the heat loss and gain terms in the steady state heat balance.

- **Energy Balance Anaerobic Digester:** The calculated values for sensible heat lost by the sludge in the digester and transferred through the walls.

- **Heat Exchange:** The quantity of heat transferred, lost and (for the gas engine) converted to mechanical energy.

- **Heat Data:** The calorific and sensible heat content of the feed, aerobic and anaerobic sludges.

- **Final Sludge Quality:** The remaining biodegradable VS concentration ($kg/(m^3)$), the total oxygen demand ($g(O_2)/kg(TSS)$), and the peak ***SOUR*** ($g(O_2)/kg(TSS).h$) and time from the start of the test at which it occurs..

- **Cost Analysis:** The current estimated capital cost for each of the principle items in the process. The projected operating cost in 7½ years (as this is the mid-term of the loan period), including repayment on the capital.

Information generated by the simulation programme has been used to compile the design charts presented in Section 8.5 below.

8.4.2 Example of a Typical Output

Figure 8.8 below displays the output produced by the simulation programme for the following system:

Aerobic reactor	R_h= 1.0 days	@ 60°C	V_{reac}= 200m^3
Anaerobic digester	R_h= 10.0 days	@ 35°C	V_{dig}= 2000m^3

Oxygenation with pure oxygen. Interstage heat exchanger installed.

```
SYSTEM DATA: DUAL DIGESTION PLANT
Feed Sludge Flow Rate of Primary Sludge  200 m3/d
Number of Streams        1      Ambient Temperature       20.0°C
Oxygenation of Aerobic Reactor: Pure Oxygen
Aerobic Reactor Size   200 m3   Anaerobic Digester Size   2000 m3
Retention Times:  Reactor   1.00 days     Digester 10.00 days
Ext Heating Units: Interstage Heat Exchange:

SLUDGE QUALITY  INFLUENT SLUDGE   REACTOR SLUDGE   DIGESTER SLUDGE
TS(kg/m3)              40.0             36.3             26.7
VS(kg/m3)              32.4             28.7             19.1
BVS(kg/m3)             19.4             15.8              3.5
Percent VS (d.m.b.)    81.0             79.1             71.6
Temperature (C)        20.0             60.0             35.0
PERCENT          AEROBIC REACTOR ANAEROBIC DIGESTER   OVERALL
VS REMOVAL             11.3             33.3             40.9

OUR (kg/m3.h) Oxygen = 0.260 Air = 0.000 Total = 0.260 (maximum = 0.756)

OXYGEN CONSUMPTION:  65.2 kg/h      1.56 T/d      571 T/annum

BIOGAS PRODUCTION:  86.0 m3/h    2064 m3/d
                    Utilised   =     0 m3/d         Wasted  =   2064 m3/d
BIOGAS QUALITY      Methane    =  55.1       Carbon Dioxide =  44.9
percent

ENERGY BALANCE AEROBIC REACTOR
Biological Heat        668       Wall Heat Loss            56
Mechanical Heat        330       Vent Gas Heat Loss        22
Recycled Heat          412       Sensible Heat Loss      1331
Total Heat In         1410       Total Heat Out          1410

ENERGY BALANCE ANAEROBIC DIGESTER
Sensible Heat Given    155       Wall Heat Loss           155

HEAT EXCHANGERS
Interstage Heat Exch.: Total Heat Available 687 = Transf.412 + Lost 275
Sludge Temp in =  60.0  Sludge Temp Out =  39.6

SLUDGE CALORIFIC VALUES (Based in BVS Concentration)
Feed Sludge   3531       Reactor Sludge   2864  Digester Sludge      999
Calorific Value Reduced by:  Oxidation     668  Biogas Production   1865

SENSIBLE HEATS (In Relation to the Feed Sludge Temperature 20.0°C)
Feed:  0 Reactor: 1347 After Heat Exchange:  660 Digester:  505

FINAL SLUDGE QUALITY
Biodegradable Volatile Solids          3.49 kgBVS/m3
The Total Oxygen Demand              254.719 gO/kgTSS
The Predicted Maximum SOUR            14.038 gO/kgTSS.h
(Occurring after   12.3 hrs of the test)

CAPITAL COST (1995)     OPERATING COSTS P.A. (2002)   (R1,000,000's)
Reactor      0.526      Oxygen           0.733
Pumps        0.184      Electrical       0.959
Air Supply   0.000      Transport        0.000
Digester     3.002      Sub Total        1.691
Boiler       0.000      Recovered        0.000
Gas Engine   0.000      Sub Total        1.691
Exchangers   0.022      Cap Repayment    0.722
Total        3.735      Total            2.413
```

Figure 8.8 Example of an Output Data Sheet from the Simulation Model

8.5 APPLICATION OF THE GENERAL SIMULATION MODEL

8.5.1 Introduction

In this section a number of applications of the general simulation model are given. Whilst the model is capable of producing a wide variety of different simulations, the following have been selected as they yield pertinent information with regard to:

- The effect of ambient (i.e. feed sludge, influent gas and surrounding air) temperature on system VS removal, biogas production and the required oxygen utilisation rate.
- The effect of increasing the proportion of external heat to the aerobic reactor on system VS removal, biogas production and the required oxygen utilisation rate.
- The effect of improved oxygen transfer efficiency of the air oxygenation device on the required rate of pure oxygen injection.
- Prediction of the minimum retention time to prevent substrate limitation, for different feed sludge types. The effects of feed sludge solids concentration and ambient temperature.

Graphical illustrations for each of the above applications are presented.

8.5.2 The Effect of Ambient Temperature on Process Performance

The effect of ambient (i.e. feed sludge, influent gas and surrounding air) temperature on the performance of two different dual digester configurations is examined (see Figure 8.9 below).

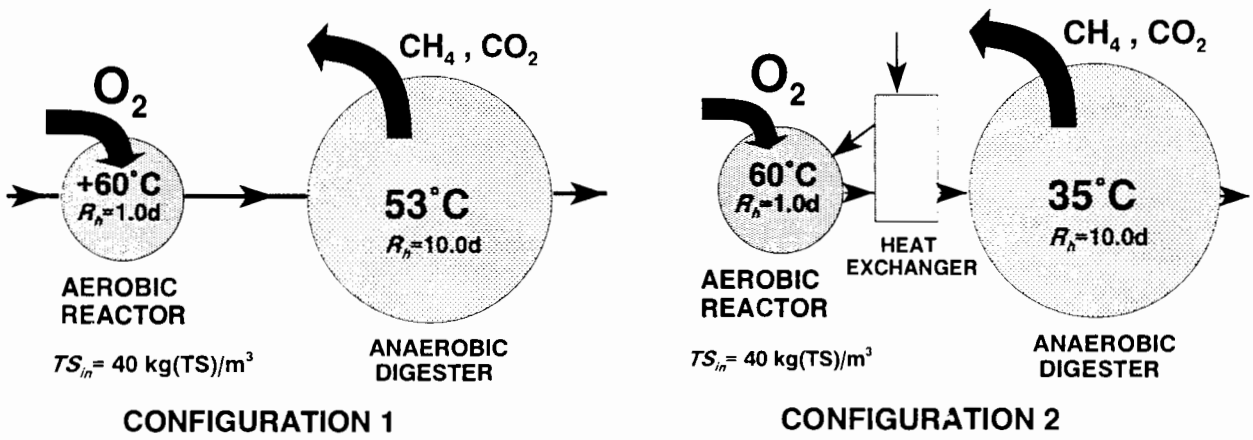


Figure 8.9 Dual Digester System Configurations N°'s 1 (thermophilic digester with no heat exchange) and 2 (mesophilic digester with interstage heat recovery), Selected to Predict the Effects of Ambient Temperature on the Primary System Parameters

In the first configuration (N°1), the anaerobic digester temperature is kept constant at 53°C (thermophilic) by controlling the aerobic reactor temperature with the pure oxygen injection rate. No external heating is provided. In the second configuration (N°2), the anaerobic digester temperature is kept constant at 35°C (mesophilic) by controlling the contact time through an interstage heat exchanger (the aerobic reactor temperature is kept constant at 60°C). In both systems, the reactor retention time is 1 day and the digester retention time 10 days. Oxygenation is with pure oxygen only. The feed sludge is primary sludge with a total solids concentration of 40 kg(TS)/m^3. The effect of ambient temperature on three of the primary process parameters; volatile solids removal efficiency, biogas production and the required aerobic reactor oxygen utilisation rate, for both configurations is shown graphically in Figures 8.10 and 8.11 below.

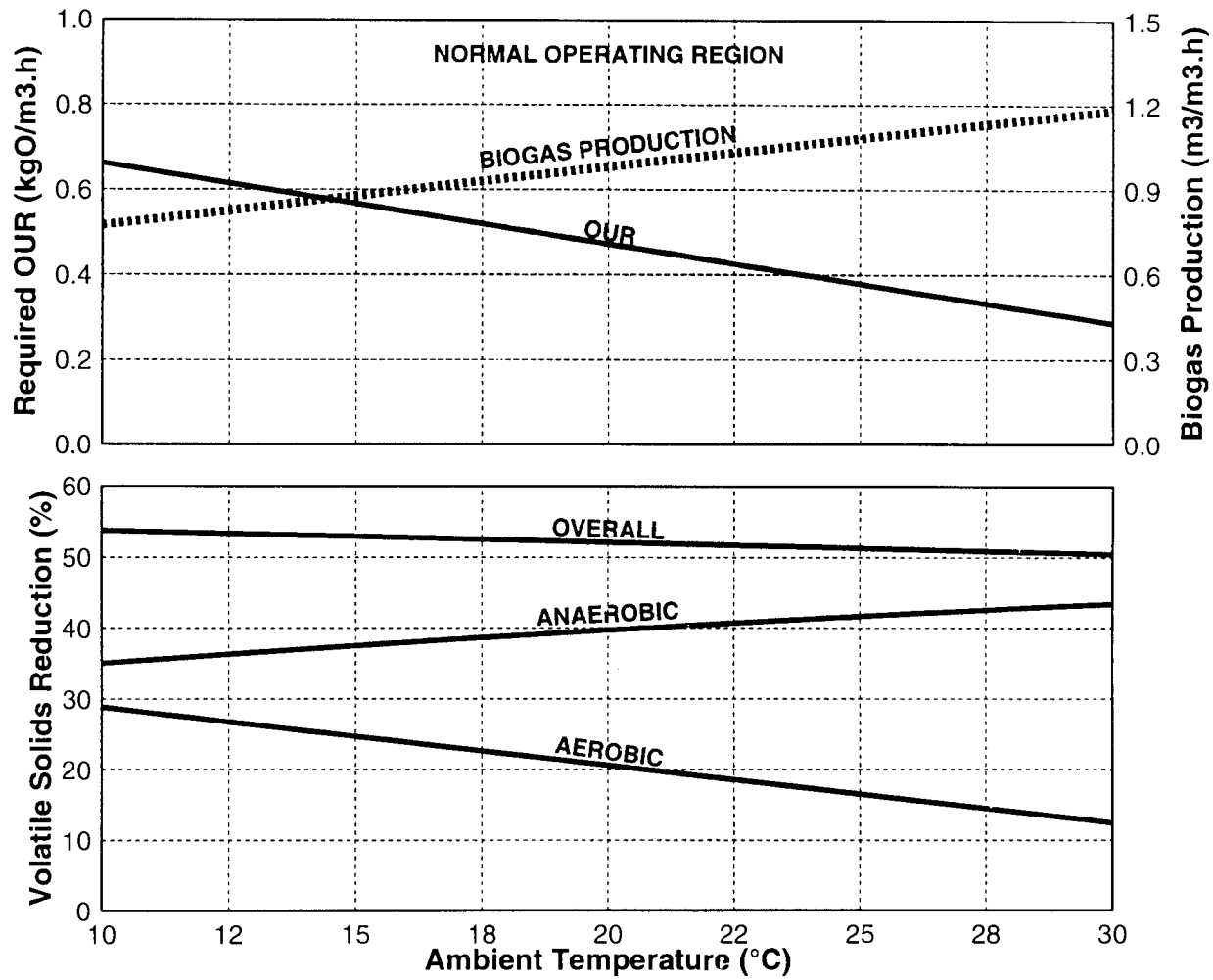


Figure 8.10 The Effect of Ambient Temperature on Volatile Solids Removal, Biogas Production and the Required Oxygen Utilisation Rate for Dual Digester System Configuration N°1 (thermophilic digester with no heat exchange; refer Figure 8.9).

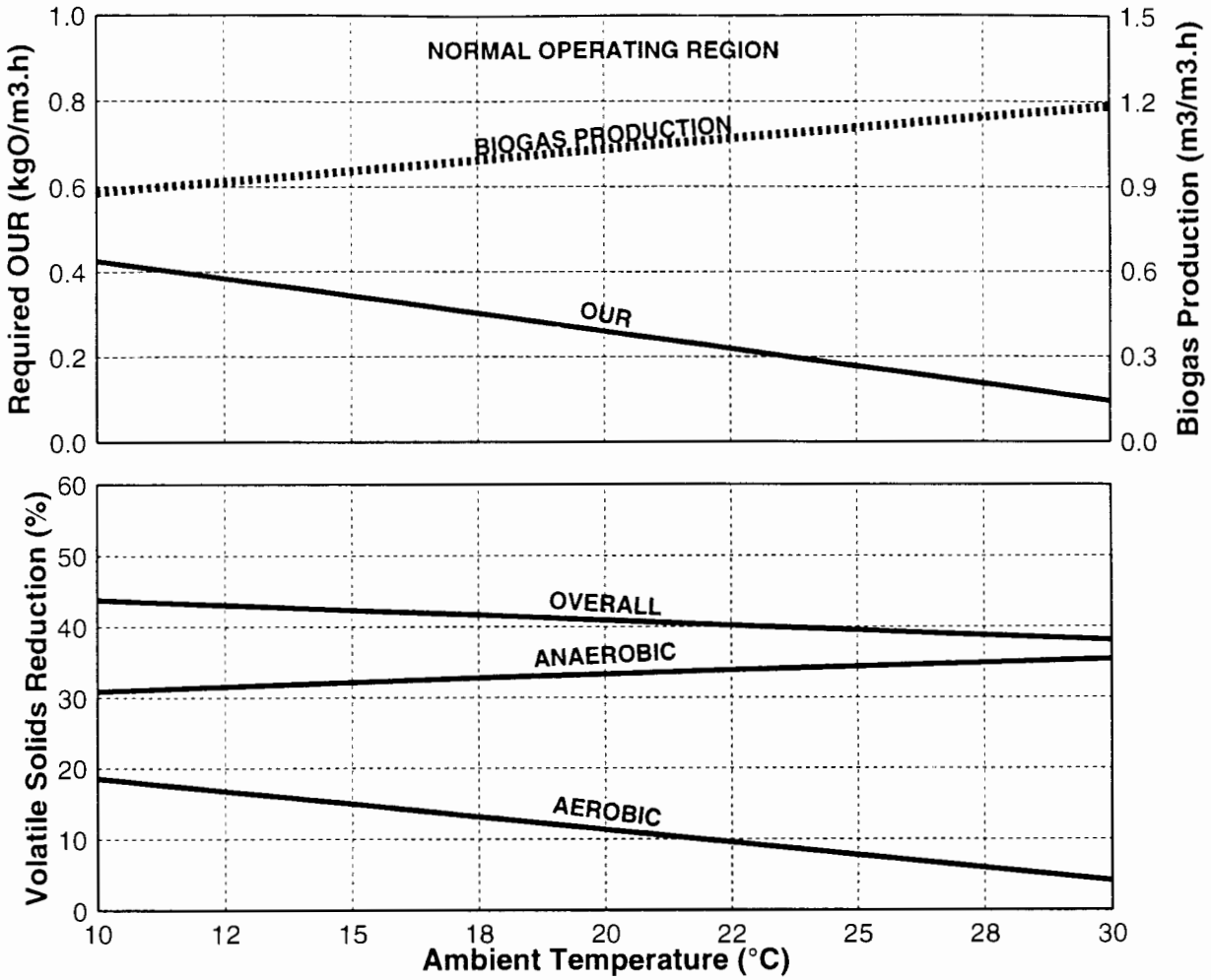


Figure 8.11 The Effect of Ambient Temperature on Volatile Solids Removal, Biogas Production and the Required Oxygen Utilisation Rate for Dual Digester System Configuration N°2 (mesophilic digester with interstage heat exchange; refer Figure 8.9).

In both instances, as the ambient temperature increases, the required oxygen utilisation rate OUR^{req} in the aerobic reactor decreases (as less biological heating is required). The reduced level of oxidation in the aerobic reactor at the warmer temperatures (shown also by a decline in the %VS removal in the reactor) leads to increased biogas production and %VS removal in the digester. However because the VS removal rate is higher under aerobic conditions than under anaerobic conditions, the overall %VS removal for both configurations shows a gradual decline as the ambient temperature increases. In the case of configuration N°2 (mesophilic digester with interstage heat exchange), the %VS removal reduces to close to the 38% limit (set for sludge stability) at the higher temperatures (25-30°C), leaving little room for system flexibility. Configuration N°1 (thermophilic digester with no heat exchange) has a much higher overall %VS removal (54% to 50%) than configuration N°2 (43% to 38%) due to the higher anaerobic digester temperature and so produces a more stable final sludge product.

8.5.3 The Effect of Improved 'Air' Transfer Efficiency on Process Performance

The effect of installing an air oxygenation device, which can produce an oxygen transfer efficiency up to 100%, on reducing the required oxygen utilisation rate attributable to the pure oxygen OUR^{O2} is graphically illustrated in Figure 8.12 below. The dual digestion plant is set up as shown by Configuration N°1 (thermophilic digester) in Figure 8.9 above. The air flow rate to the reactor is at 375m³(STP)/h (the calculated air flow rate which reduces the OUR^{O2} to zero at an 'air' transfer efficiency of 100%).

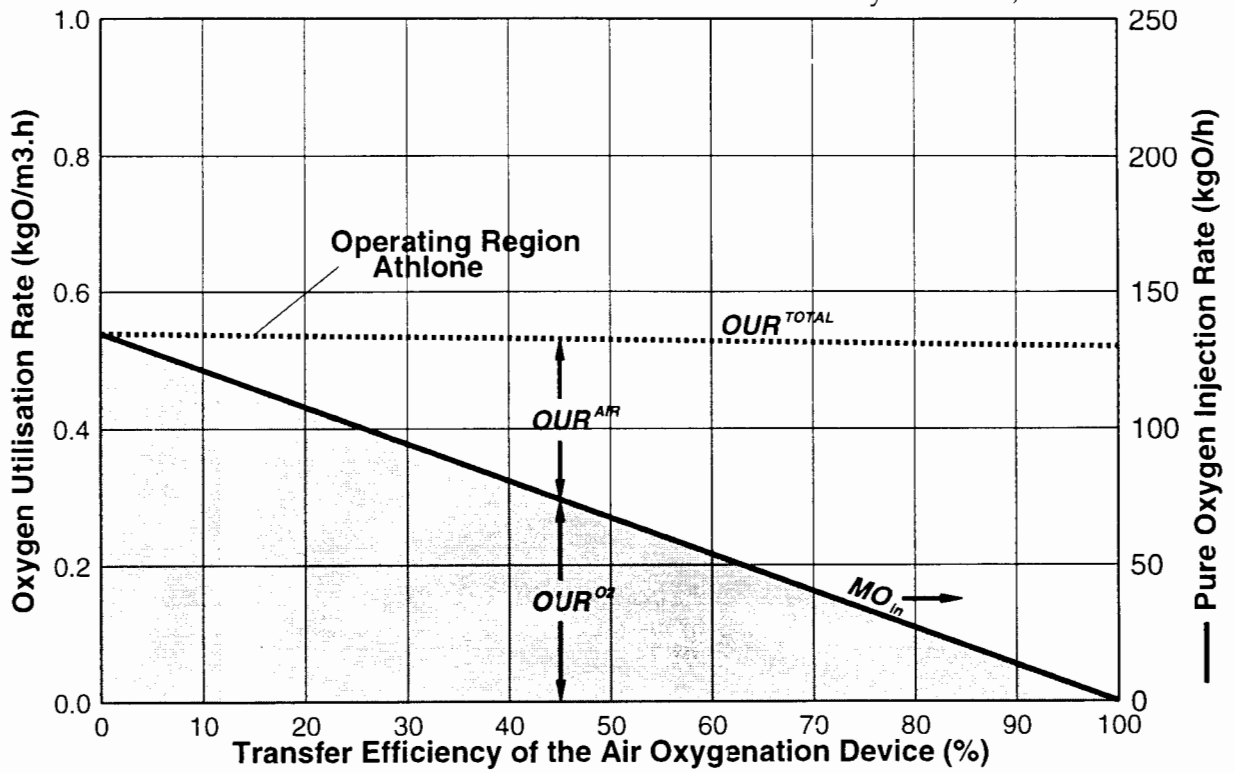


Figure 8.12 The Effect of Improved 'Air' Transfer Efficiency on the Required Pure Oxygen Utilisation Rate OUR^{O2}

Figure 8.12 shows that it is possible to maintain a reactor retention time of 1 day at high 'air' oxygen transfer efficiencies. It had been one of the objectives of phase II (refer Section 6.3), to exploit the foaming phenomenon by obtaining improved 'air' transfer efficiencies during foaming and using pure oxygen injection (supplementation) to provide the reactor temperature control, which was lacking during phase I when oxygenation was with air alone. Unfortunately, foaming did not occur during phase II[1]. The type of air oxygenation device employed at Athlone (course bubble diffusion) is considered far from ideal for this type of application. A more appropriate method would be to inject the air into the recirculation line, under pressure, as is the case for the pure

[1] The exact reasons for the lack of foaming during phase II could not be established. However, it was speculated that the reduction in recirculation line flow rate (from >1000m³/h to 700m³/h) may have been a contributing factor.

oxygen; a method employed by Wolinski (1985) who reported 'air' oxygen transfer efficiencies close to 100% during foaming. Obviously, it would be impractical to simultaneously inject air and pure oxygen at the same point.

8.5.4 The Effect of External Reactor Heating on Process Performance

The effect of supplementing the biological heating rate in the aerobic reactor by external heating of the cold reactor influent feed sludge was simulated using the system configuration (N°3) depicted in Figure 8.13. External heat is provided by burning biogas in a conventional hot water boiler and transferring the heat generated to the cold influent feed sludge via a heat exchanger (refer Section 8.3.9 above). In the simulation the heat source for the reactor was varied from 0% external (all biological) to 100% external (zero biological). The effect on the primary process parameters, %VS removal, biogas production and the reactor oxygen utilisation rate is illustrated in Figure 8.14 below.

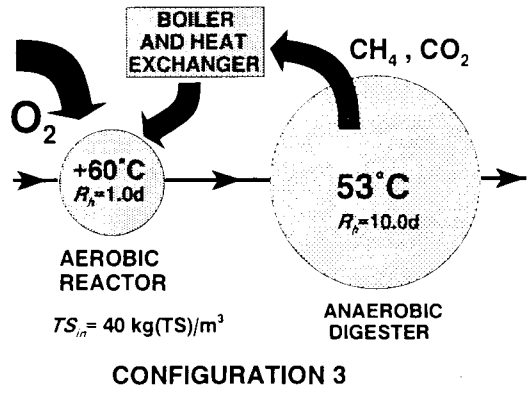


Figure 8.13 Dual Digester System Configuration N°3: Thermophilic digester with afterstage heat exchange

At 0% external heat, the system runs as a 'pure' dual digestion system, with all the aerobic reactor heat requirements provided by biological heating and the input of energy from the mixing device. With the proportion of external heat at 100%, the system is equivalent to conventional (in this instance thermophilic) digestion with pre-pasteurisation. Comparison between the two extremes, shows that (in this example) dual digestion can reduce biogas production in the digester by as much as 34%. However, in comparison with thermophilic digestion, 64% of the biogas produced is required to produce the necessary heat to maintain thermophilic temperatures.

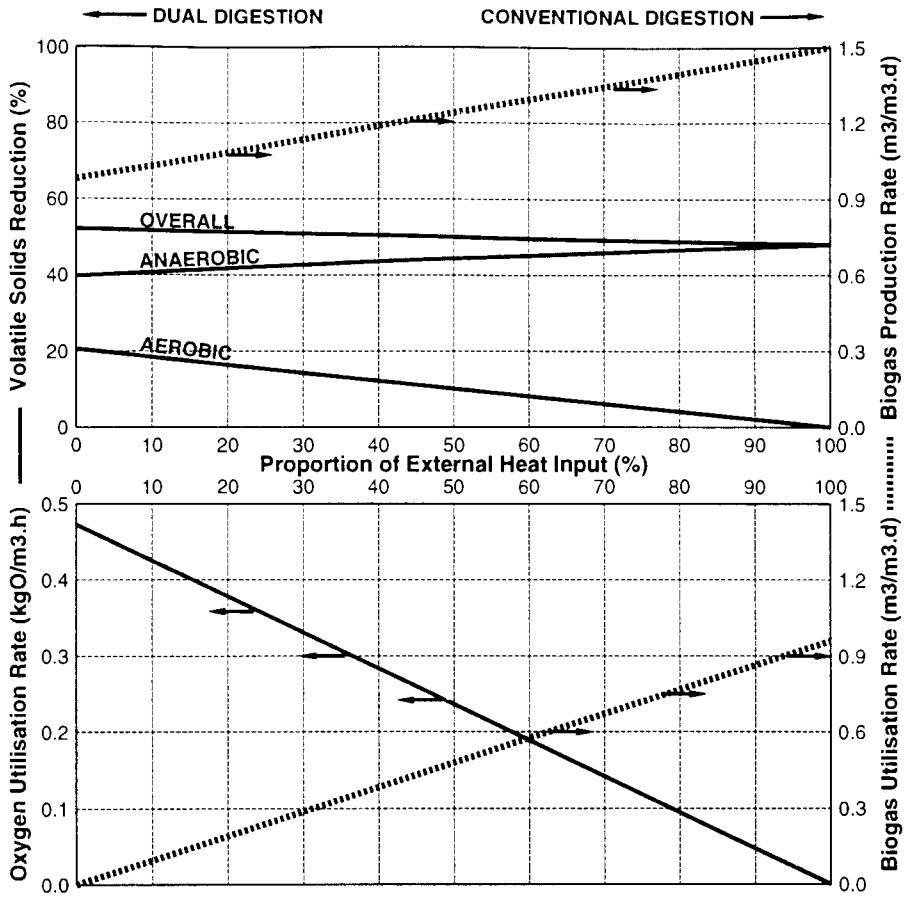


Figure 8.14 The Effect of External Heating on Volatile Solids Removal, Biogas Production and the Required Oxygen Utilisation Rate for Dual Digester System Configuration N°3 (thermophilic digester with afterstage heat exchange; refer Figure 8.13).

8.5.5 The Prediction of the Onset of Substrate Limitation

In the operation of the aerobic reactor it is essential that the reactor be operated under *oxygen limiting* conditions. Under such conditions, the biological oxygen utilisation rate ***OUR*** will be fixed by the oxygen transfer rate ***OTR*** of the oxygenation device(s) (***OUR*** = ***OTR)*** and there will be sufficient substrate (in the form of readily biodegradable volatile solids; RBVS) to support the biological oxidation reactions. However, should the concentration of RBVS fall below the required level, then the reactor will become *substrate limited.* Under such conditions the ***OUR*** will fall below the required ***OTR*** and the performance of the aerobic reactor will be adversely affected, i.e. there will be a decline in temperature due to a reduction in the biological heating rate (which is linked to the ***OUR***) and a loss of temperature control via the oxygen supply rate. It is essential therefore that in the design and/or operation of the aerobic reactor, consideration must be given to the factors which can contribute to substrate limitation. At short retention times (1 to 3 days) using pure oxygen, the factors which have an affect on substrate limitation are:

- Aerobic reactor retention time
- Sludge type (i.e. primary, waste activated etc.)
- Feed sludge total solids concentration
- Ambient (i.e. feed sludge, influent gas and surrounding air) temperature

The effect of these factors on the retention time at which the reactor becomes substrate limited is considered in three examples below. The retention time at which substrate limitation commences, can be considered to be the minimum retention time required for operation under oxygen limiting conditions R_h^{min} ; below R_h^{min} the reactor will be substrate limited. Since the actual growth kinetics of the thermophilic organisms are not considered in the simulation model, it is accepted that at retention times below 0.8d, the organism will commence to be washed out of the aerobic reactor due to maximum specific growth rate limitation. This will cause a reduction in ***OUR*** for reasons other than substrate or oxygen limitation. In the simulation, the dual digestion system is assumed to be operating as configuration N°1 (thermophilic digestion with no heat exchange; refer Figure 8.9).

Example 1: Sludge Type

In the first example (Figure 8.15), the maximum oxygen utilisation rates OUR^{max} for different sludge types at a feed concentration of 40 kg(TS)/m^3 (dotted lines) and the required oxygen utilisation rate OUR^{req} for operation at a certain retention time (solid line) are plotted against aerobic reactor retention time. The sludge types examined are as follows (data regarding the VS and BVS fractions for each sludge type are contained in Table 8.1 above):

- Primary sludge
- Primary/humus sludge mixture
- Primary/waste activated sludge mixture
- Waste activated sludge

In order for the reactor to remain under oxygen limiting conditions, OUR^{max} must be greater than OUR^{req}. The minimum required retention time for operation R_h^{min} is the point at which the OUR^{max} and OUR^{req} lines intersect. At retention times below R_h^{min} the ***OUR*** will fall below OUR^{req}, the reactor will become substrate limited and the sludge temperature will decrease because OUR^{req} is not limited at OUR^{max}.

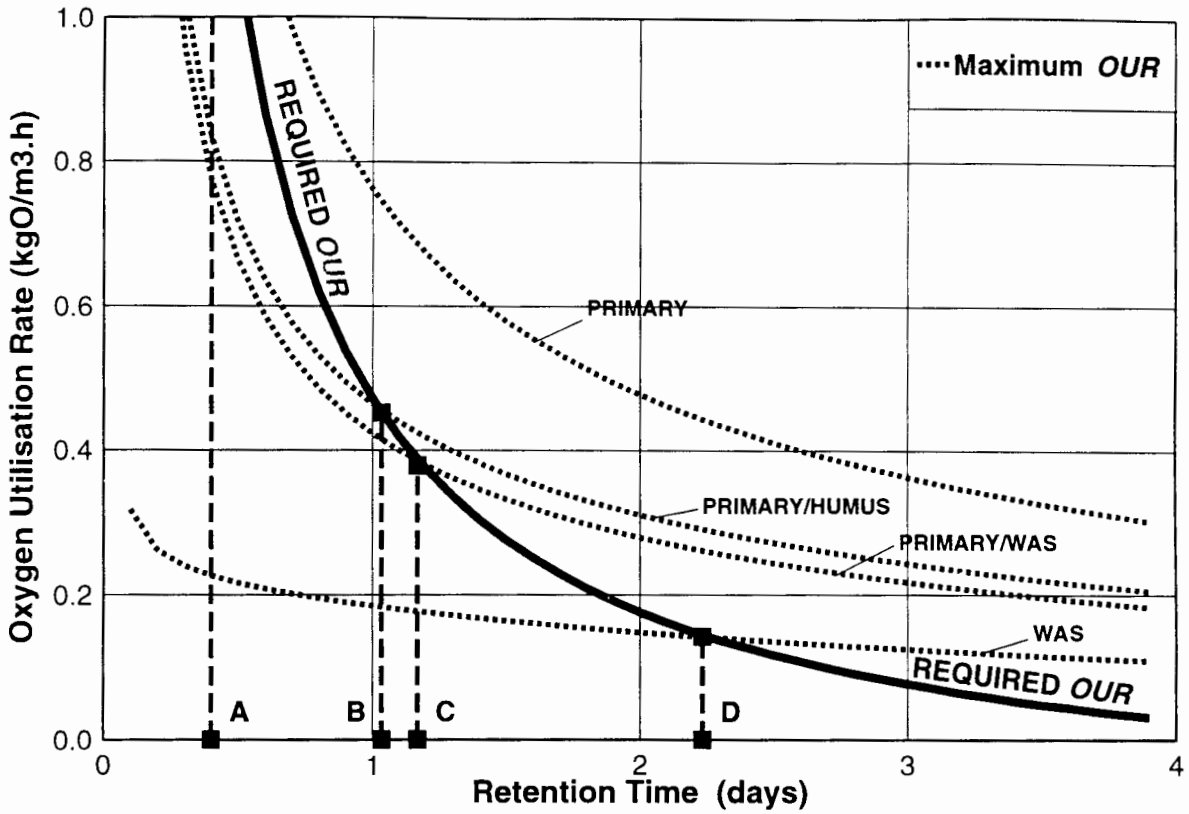


Figure 8.15 **The Variation in Maximum Oxygen Utilisation Rate OUR^{max} for Different Sludge Types (dotted lines) and Required Oxygen Utilisation Rate OUR^{req} (solid line) ($kg(O_2)/m^3.h$) with Aerobic Reactor Retention Time. Oxygenation is with Pure Oxygen. Feed Sludge Total Solids Concentration = 40 $kg(TS)/m^3$. Ambient Temperature = 20°C.**

From Figure 8.15, the estimated minimum operational retention times R_h^{min} at which it is possible to operate under oxygen limiting conditions (reactor temperature of 60°C and ambient temperature of 20°C) for the different sludge types are given in Table 8.2.

Table 8.2 **The Minimum Possible Aerobic Reactor Retention Times for Oxygen Limitation Operation for Different Sludge Types. Oxygenation with Pure Oxygen. Ambient Temperature = 20°C**

Sludge Type	Minimum Retention Time R_h^{min} (days)	Ref. in Fig 8.15
Primary	0.40†	Point A
Primary/Humus	1.05	Point B
Primary/Waste Activated	1.17	Point C
Waste Activated	2.23	Point D

† The aerobic bacteria may be washed out of the reactor, due to maximum specific growth rate limitations, before this retention time is achieved.

Example 2: Feed Sludge Concentration

In the first example, the feed sludge total solids concentration was fixed at 40 kg(TS)/m^3 and the minimum retention R_h^{min} was predicted. In the second example, the retention time is fixed at 1.0 day, and the minimum required feed sludge total solids concentration is predicted to ensure an OUR^{max} equal to or greater than OUR^{req} i.e. to maintain oxygen limiting conditions. In Figure 8.16, the required oxygen utilisation rate OUR^{req} for operation at 1 day is shown as a horizontal solid line at an OUR of 0.472 kg(O_2)/m^3.h. The maximum oxygen utilisation rate OUR^{max} possible, versus solids concentrations, for each sludge type are shown as dotted lines.

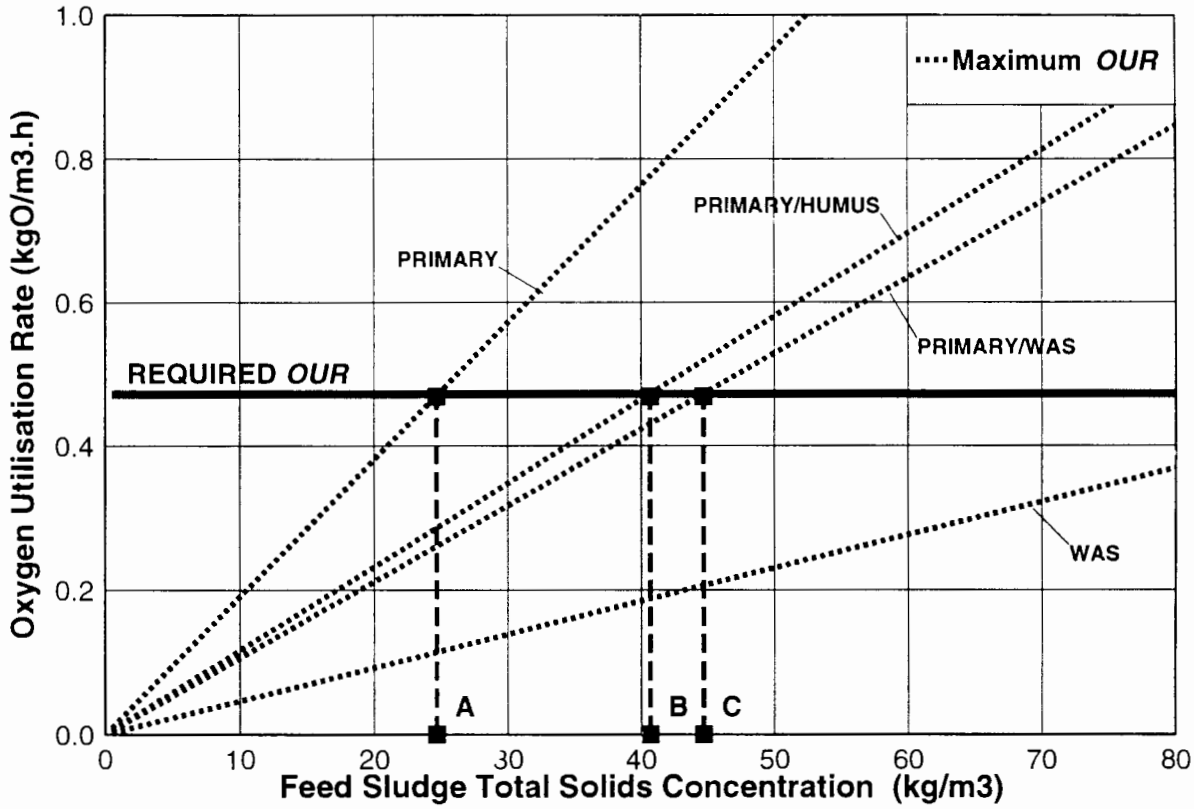


Figure 8.16 The Variation in the Maximum Possible Oxygen Utilisation Rate (for different sludge types) with Feed Sludge Total Solids Concentration. (pure oxygen system, reactor retention time of 1day).

The minimum required feed sludge total solids concentration TS_{in}^{min} for operation at 1 day (under the conditions specified above) for each sludge type, is determined from the points of intersection of the OUR^{max} (dotted) lines and the horizontal OUR^{req} (solid) line in Figure 8.16 above. The values obtained for TS_{in}^{min} for the four different sludge types are given in Table 8.3 below.

Whilst it is practically possible to meet the TS_{in}^{min} feed concentrations listed in Table 8.3 for each of the first three sludges which consist of or contain significant amounts of primary sludge, the waste activated sludge contains insufficient energy, even at a

Table 8.3 **The Minimum Required Feed Sludge Total Solids Concentrations for Different Sludge Types. Oxygenation with Pure Oxygen.**

Sludge Type	Minimum Feed Sludge Concentration TS_{in}^{min} (kg(TS)/m^3)	Ref. in Fig 8.16
Primary	24.5	Point A
Primary/Humus	41.0	Point B
Primary/Waste Activated	44.5	Point C
Waste Activated	>80.0	-

concentration of 80 kg(TS)/m^3, for the OUR^{max} to exceed the OUR^{req}. Operation at a retention time of 1 day is therefore not possible for waste activated sludge. At 2 days retention time, the OUR^{req} is about half that at 1 day (i.e. about 0.236 kg(O_2)/m^3.h). The required influent solids concentration (from Figure 8.16) would be around 50 kg(TS)/m^3.

Example 3: Affect of Ambient Temperature

The effect of ambient (i.e. feed sludge, influent gas and surrounding air) temperature on the minimum reactor retention time R_h^{min} (to ensure $OUR^{max} \geq OUR^{req}$) is illustrated in Figure 8.17 below. Here, plots for both waste activated sludge and a primary/humus sludge mixture are presented. As in the previous two examples, oxygenation is with pure oxygen. The feed sludge solids concentration is assumed to be at 40kg(TS)/m^3.

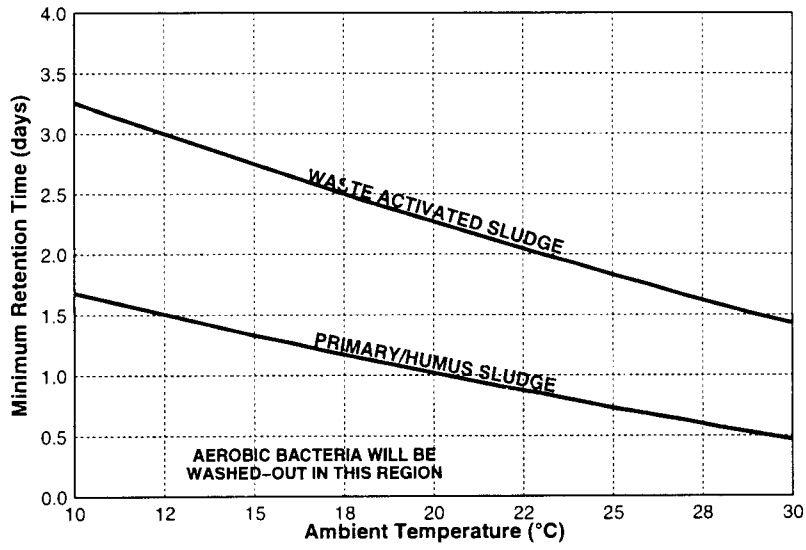


Figure 8.17 **Effect of Ambient Temperature on the Minimum Required Aerobic Retention Time. Oxygenation with Pure Oxygen. Feed Sludge Solids Concentration at 40kg(TS)/m^3.**

8.5.6 Comparison Between Different Sludge Treatment System Configurations

In this Section, the general dual digestion simulation model is used to compare different dual digester systems with that of conventional anaerobic digestion (both mesophilic and thermophilic). The objective is to evaluate the feasibility of the dual digestion system compared with conventional treatment in terms of anaerobic process and product stability and in terms of total system cost. Nine sludge treatment systems (labled **A** to **I**) are compared viz.

A basic dual digestion (no external source of heating)
B dual digestion with interstage heat exchange
C dual digestion with afterstage heat exchange
D dual digestion with supplementary heat from a conventional boiler
E dual digestion with supplementary heat from a gas engine
F mesophilic digestion: heating with a conventional boiler
G mesophilic digestion: using the heat from a gas engine
H thermophilic digestion: heating with a conventional boiler
I thermophilic digestion: using the heat from a gas engine

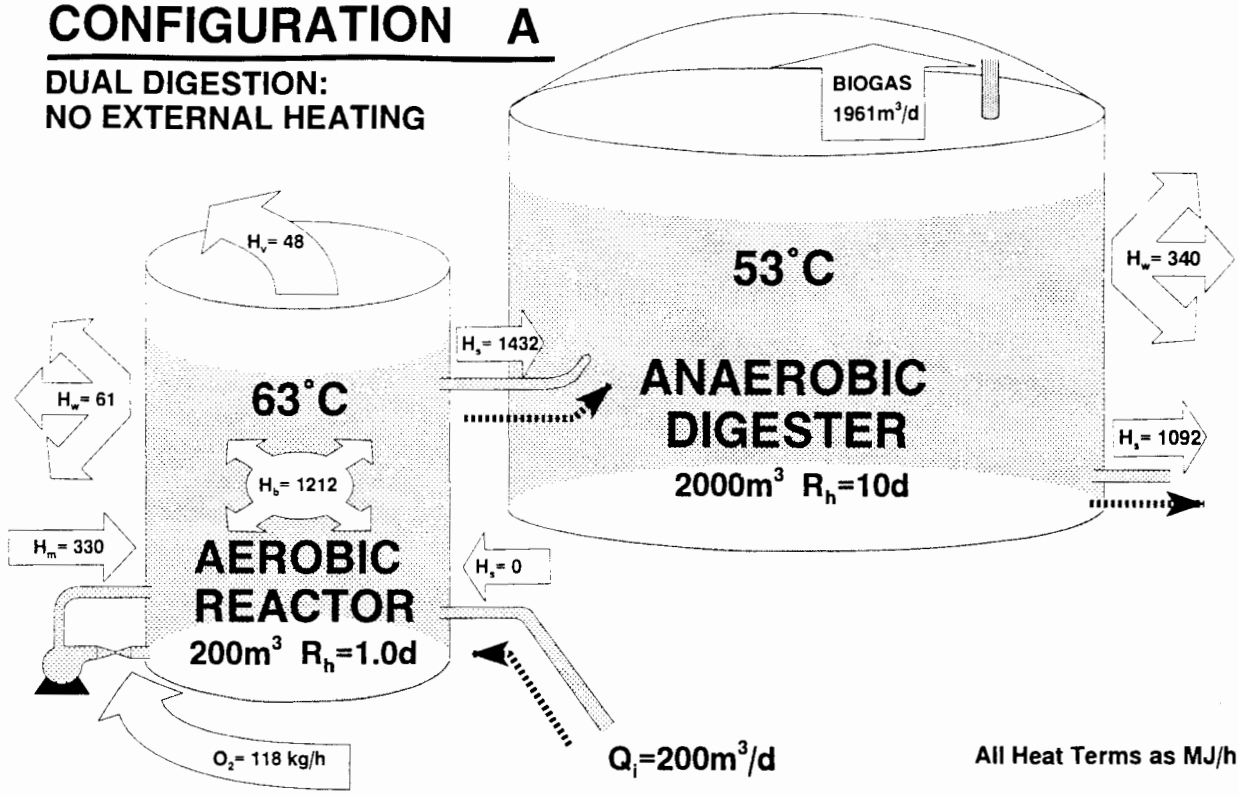
In each system, 200m^3/d of primary sewage sludge (4%TS) is treated, with the average ambient temperature at 20°C (typical of South African conditions). In each dual digester system (**A** to **E**), oxygenation is with pure oxygen alone. The aerobic reactor retention time is at one day and the anaerobic digester retention time at ten days. For conventional mesophilic digestion (**F** and **G**), the digester retention time is set at twenty days. For conventional thermophilic digestion (**H** and **I**), the digester retention time is set at ten days.

In determining the overall cost for each system, the repayment on the capital cost is included in the total operating cost (SAR 1,000,000's p.a. 2002) (see Section 8.3.10 above for details). It is recognised that the most serious mechanical problem of the aerobic system is the mechanical wear and tear of the sludge recirculation pump and in each case a stand-by pump has been budgeted for. In quoting the digester upgrade cost, it is presumed that the only item of existing equipment is the anaerobic digester (i.e. without heating equipment).

The results of the simulation of each system configuration (**A** to **I**) are presented on the nine following pages. A schematic of each plant is provided followed by a tabulation of all the relevant operating data, afterwhich there is a discussion on the comparative merits of each system. A summary of the results is presented in Table 8.4 below

CONFIGURATION A

DUAL DIGESTION:
NO EXTERNAL HEATING



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	33.3	23.1
Volatile Solids	kg(VS)/m³	32.4	25.7	15.5
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	12.8	0.3
%Volatile Solids	dry mass basis	81.0	77.2	67.1
Removal Efficiency			**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%		20.6	39.7	52.1

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	459
Destroyed via Pure Oxygen Utilisation	94
Load on Anaerobic Digester	365
Destroyed via Methane Production	145
Final Sludge	220

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	1961
Utilised	%	0
Wasted	%	100
Methane Concentration	%	61.8
Carbon Dioxide Concentration	%	38.2

4: PURE OXYGEN SUPPLIED

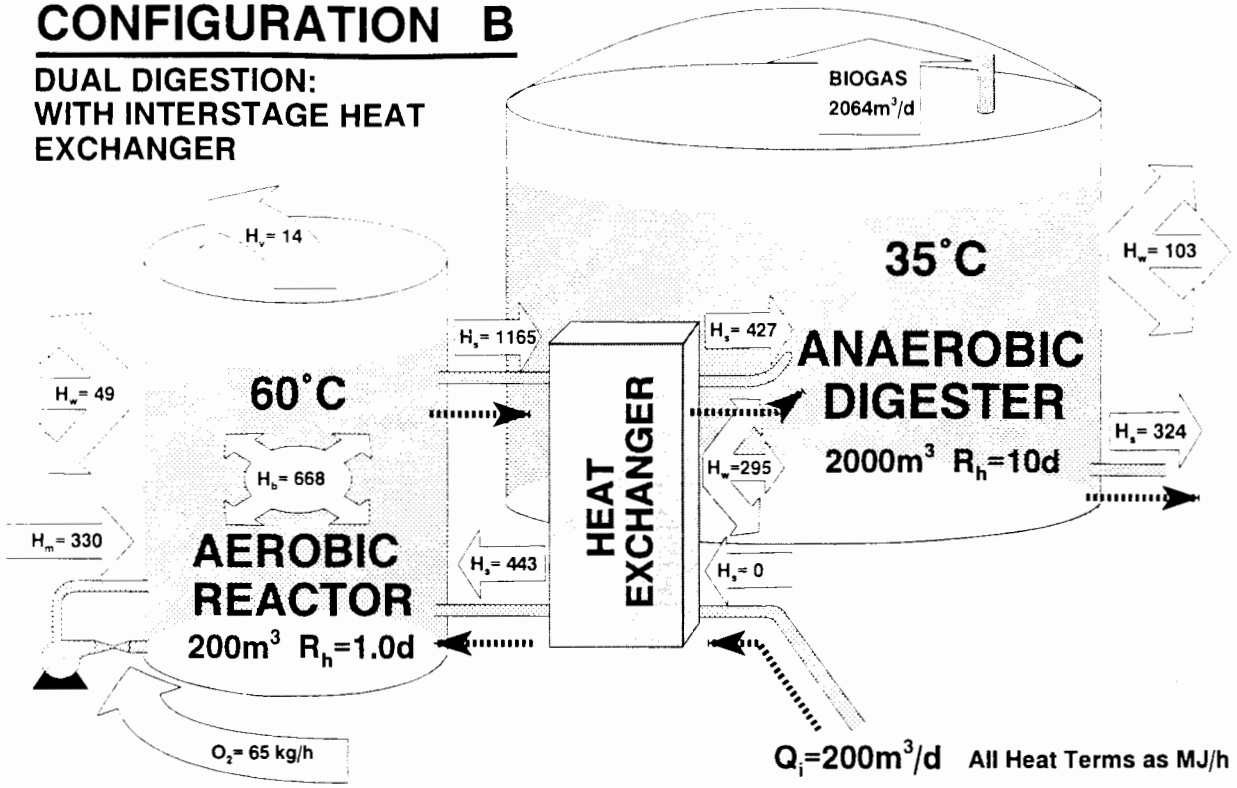
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	118
" "	t(O_2)/d @ 20°C	2.84
Summer	" " @ 25°C	2.28
Winter	" " @ 15°C	3.40

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	3.71
Operating (inc. Cap. Repayment)	2002	2.74
Upgrade Capital	1995	0.71
Operating (inc. Cap. Repayment)	2002	2.16

CONFIGURATION B

DUAL DIGESTION: WITH INTERSTAGE HEAT EXCHANGER



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	36.3	26.7
Volatile Solids	kg(VS)/m³	32.4	28.7	19.1
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	15.8	3.5
%Volatile Solids	dry mass basis	81.0	79.1	71.6
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	11.3	33.3	40.9

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	459
Destroyed via Pure Oxygen Utilisation	52
Load on Anaerobic Digester	407
Destroyed via Methane Production	136
Final Sludge	271

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2064
Utilised	%	0
Wasted	%	100
Methane Concentration	%	55.1
Carbon Dioxide Concentration	%	44.9

4: PURE OXYGEN SUPPLIED

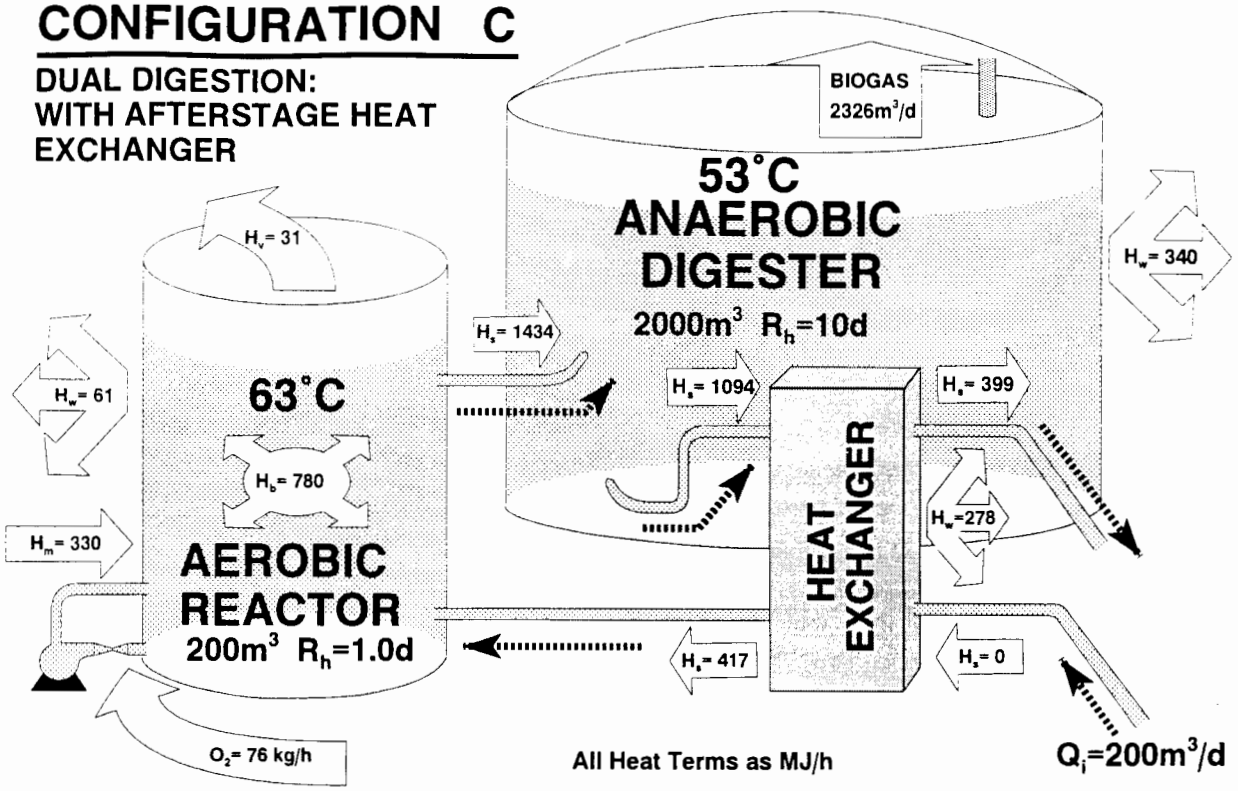
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	65
" "	t(O_2)/d @ 20°C	1.56
Summer	" " @ 25°C	1.07
Winter	" " @ 15°C	2.06

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	3.74
Operating (inc. Cap. Repayment)	2002	2.18
Upgrade Capital	1995	0.74
Operating (inc. Cap. Repayment)	2002	1.60

CONFIGURATION C

DUAL DIGESTION: WITH AFTERSTAGE HEAT EXCHANGER



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	35.7	23.6
Volatile Solids	kg(VS)/m³	32.4	28.1	16.0
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	15.1	0.4
%Volatile Solids	dry mass basis	81.0	78.7	67.8
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	13.2	43.1	50.7

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	459
Destroyed via Pure Oxygen Utilisation	61
Load on Anaerobic Digester	398
Destroyed via Methane Production	171
Final Sludge	227

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2326
Utilised	%	0
Wasted	%	100
Methane Concentration	%	61.8
Carbon Dioxide Concentration	%	38.2

4: PURE OXYGEN SUPPLIED

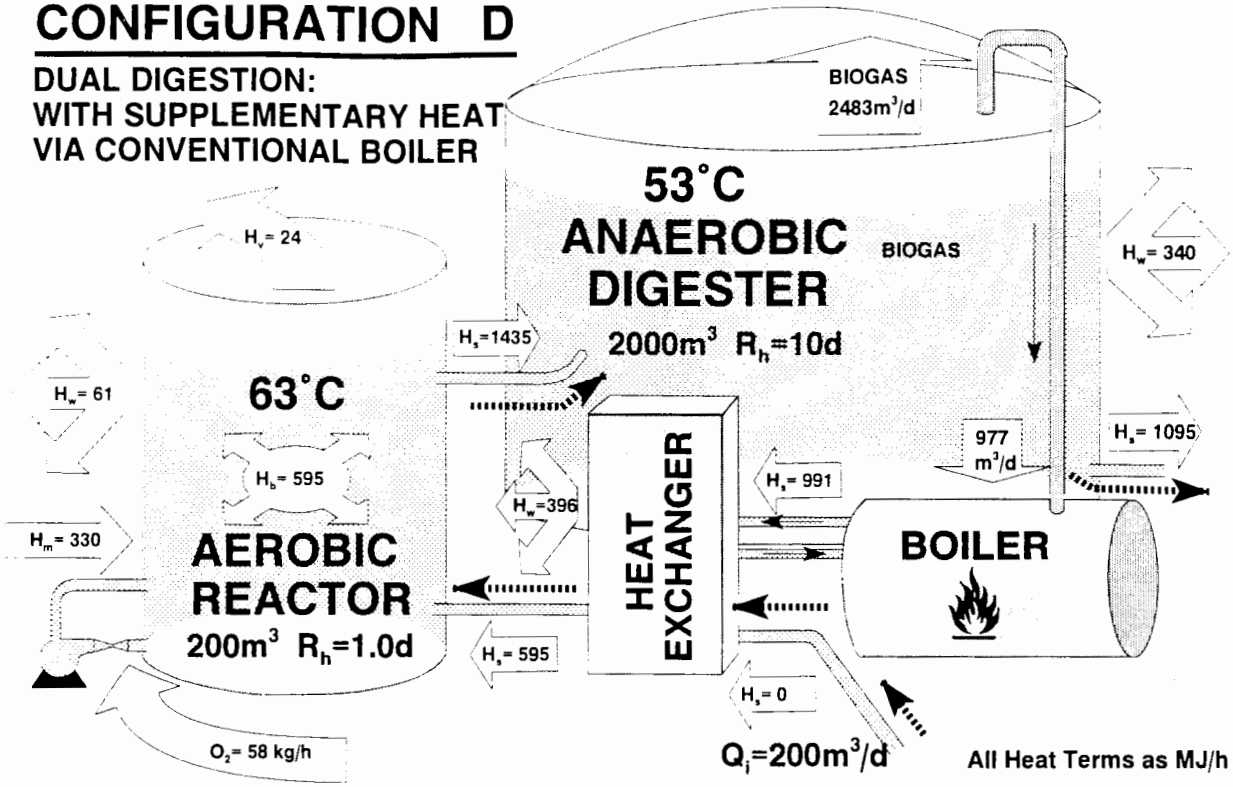
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	76
" "	t(O_2)/d @ 20°C	1.83
Summer	" " @ 25°C	1.42
Winter	" " @ 15°C	2.23

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	3.74
Operating (inc. Cap. Repayment)	2002	2.54
Upgrade Capital	1995	0.73
Operating (inc. Cap. Repayment)	2002	1.96

CONFIGURATION D

DUAL DIGESTION:
WITH SUPPLEMENTARY HEAT
VIA CONVENTIONAL BOILER



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	36.7	23.8
Volatile Solids	kg(VS)/m³	32.4	29.1	16.2
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	16.2	0.4
%Volatile Solids	dry mass basis	81.0	79.3	68.1
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	10.1	44.4	50.0

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	459
Destroyed via Pure Oxygen Utilisation	47
Load on Anaerobic Digester	412
Destroyed via Methane Production	182
Final Sludge	230

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2483
Utilised	%	39
Wasted	%	61
Methane Concentration	%	61.8
Carbon Dioxide Concentration	%	38.2

4: PURE OXYGEN SUPPLIED

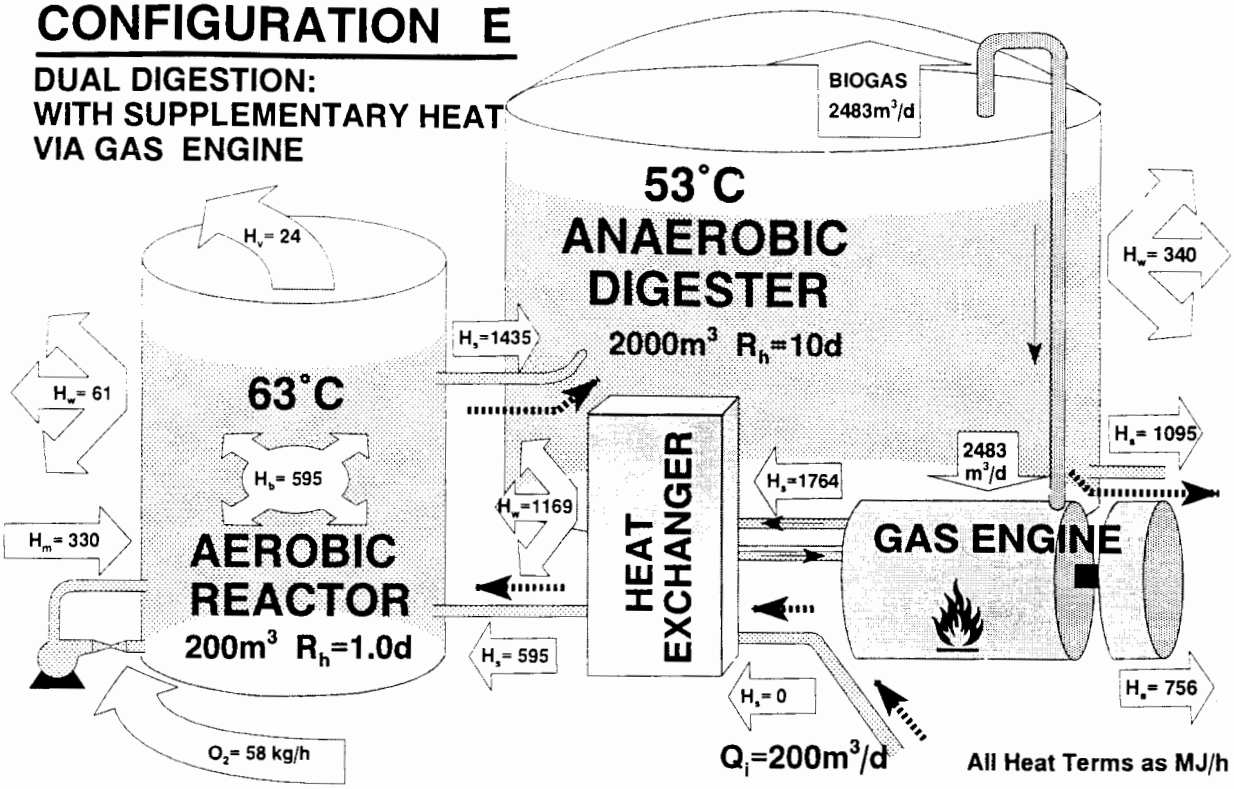
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	58
" "	t(O_2)/d @ 20°C	1.39
Summer	" " @ 25°C	1.12
Winter	" " @ 15°C	1.67

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	4.23
Operating (inc. Cap. Repayment)	2002	2.43
Upgrade Capital	1995	1.23
Operating (inc. Cap. Repayment)	2002	1.85

CONFIGURATION E

DUAL DIGESTION:
WITH SUPPLEMENTARY HEAT
VIA GAS ENGINE



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	36.7	23.8
Volatile Solids	kg(VS)/m³	32.4	29.1	16.2
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	16.2	0.4
%Volatile Solids	dry mass basis	81.0	79.3	68.1
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	10.1	44.4	50.0

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	459
Destroyed via Pure Oxygen Utilisation	47
Load on Anaerobic Digester	412
Destroyed via Methane Production	182
Final Sludge	230

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2483
Utilised	%	100
Wasted	%	0
Methane Concentration	%	61.8
Carbon Dioxide Concentration	%	38.2

4: PURE OXYGEN SUPPLIED

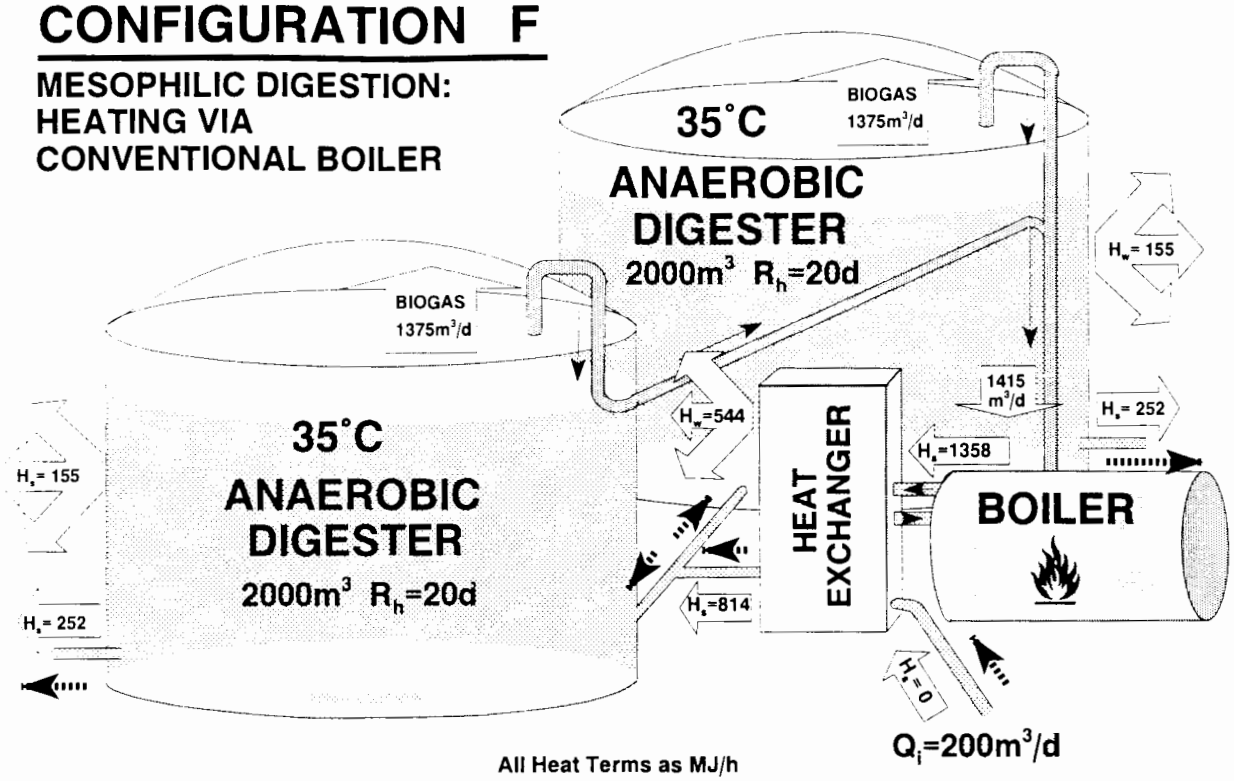
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	58
" "	t(O_2)/d @ 20°C	1.39
Summer	" " @ 25°C	1.12
Winter	" " @ 15°C	1.67

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	6.25
Operating (inc. Cap. Repayment)	2002	1.77
Upgrade Capital	1995	3.24
Operating (inc. Cap. Repayment)	2002	1.19

CONFIGURATION F

MESOPHILIC DIGESTION: HEATING VIA CONVENTIONAL BOILER



All Heat Terms as MJ/h

1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	-	26.4
Volatile Solids	kg(VS)/m³	32.4	-	18.8
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	-	2.5
%Volatile Solids	dry mass basis	81.0	-	71.3
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	-	41.8	41.8

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	-
Destroyed via Pure Oxygen Utilisation	-
Load on Anaerobic Digester	459
Destroyed via Methane Production	193
Final Sludge	266

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2750
Utilised	%	51
Wasted	%	49
Methane Concentration	%	58.5
Carbon Dioxide Concentration	%	41.5

4: PURE OXYGEN SUPPLIED

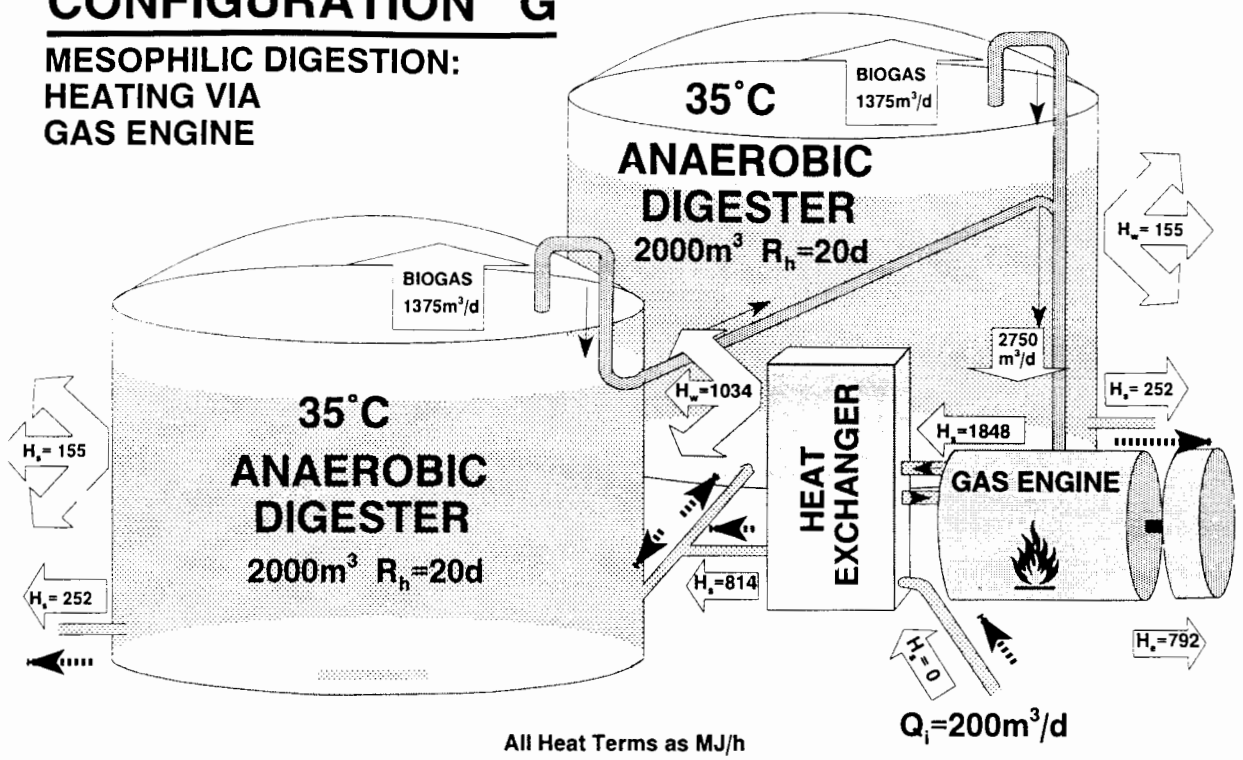
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	-
" "	t(O_2)/d @ 20°C	-
Summer	" " @ 25°C	-
Winter	" " @ 15°C	-

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	7.46
Operating (inc. Cap. Repayment)	2002	2.62
Upgrade Capital	1995	1.42
Operating (inc. Cap. Repayment)	2002	1.46

CONFIGURATION G

MESOPHILIC DIGESTION: HEATING VIA GAS ENGINE



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	-	26.4
Volatile Solids	kg(VS)/m³	32.4	-	18.8
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	-	2.5
%Volatile Solids	dry mass basis	81.0	-	71.3
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	-	41.8	41.8

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	-
Destroyed via Pure Oxygen Utilisation	-
Load on Anaerobic Digester	459
Destroyed via Methane Production	193
Final Sludge	266

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2750
Utilised	%	100
Wasted	%	0
Methane Concentration	%	58.5
Carbon Dioxide Concentration	%	41.5

4: PURE OXYGEN SUPPLIED

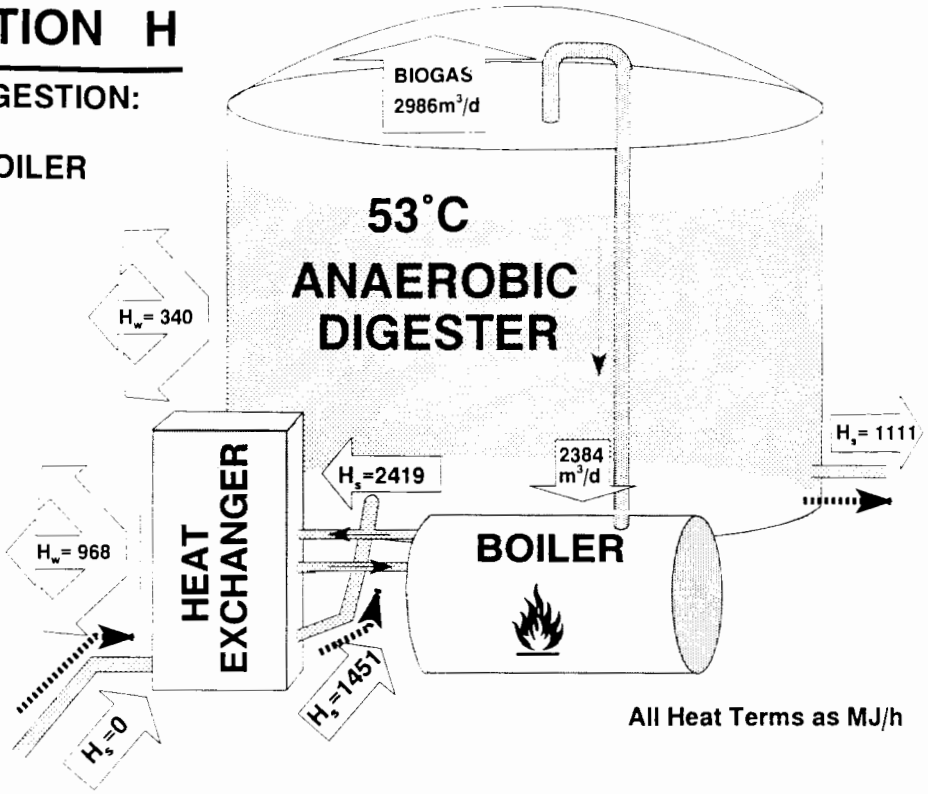
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	-
" "	t(O_2)/d @ 20°C	-
Summer	" " @ 25°C	-
Winter	" " @ 15°C	-

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	8.67
Operating (inc. Cap. Repayment)	2002	1.75
Upgrade Capital	1995	2.66
Operating (inc. Cap. Repayment)	2002	0.59

CONFIGURATION H

THERMOPHILIC DIGESTION: HEATING VIA CONVENTIONAL BOILER



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	-	24.4
Volatile Solids	kg(VS)/m³	32.4	-	16.8
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	-	0.5
%Volatile Solids	dry mass basis	81.0	-	68.9
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	-	48.0	48.0

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	-
Destroyed via Pure Oxygen Utilisation	-
Load on Anaerobic Digester	459
Destroyed via Methane Production	221
Final Sludge	238

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2986
Utilised	%	80
Wasted	%	20
Methane Concentration	%	61.8
Carbon Dioxide Concentration	%	38.2

4: PURE OXYGEN SUPPLIED

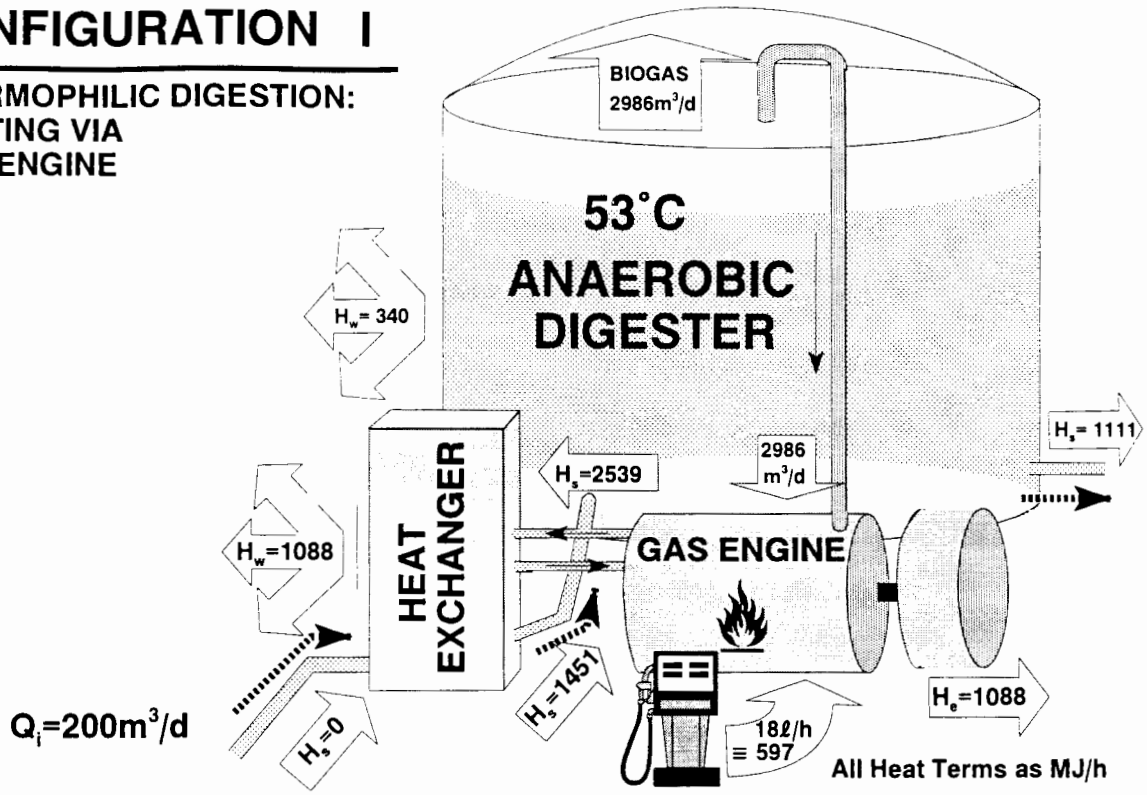
Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	-
" "	t(O_2)/d @ 20°C	-
Summer	" " @ 25°C	-
Winter	" " @ 15°C	-

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	4.28
Operating (inc. Cap. Repayment)	2002	1.33
Upgrade Capital	1995	1.27
Operating (inc. Cap. Repayment)	2002	0.74

CONFIGURATION I

THERMOPHILIC DIGESTION: HEATING VIA GAS ENGINE



1: SLUDGE QUALITY

Parameter		Raw Feed	Aerobic	Anaerobic
Total Solids	kg(TS)/m³	40.0	-	24.4
Volatile Solids	kg(VS)/m³	32.4	-	16.8
Biodegradable Volatile Solids	kg(BVS)/m³	19.4	-	0.5
%Volatile Solids	dry mass basis	81.0	-	68.9
Removal Efficiency		**Aerobic**	**Anaerobic**	**Overall**
Volatile Solids	%	-	48.0	48.0

2: COD BALANCE

kg(COD)/h	Rate
Load on Aerobic Reactor	-
Destroyed via Pure Oxygen Utilisation	-
Load on Anaerobic Digester	459
Destroyed via Methane Production	221
Final Sludge	238

3 BIOGAS CHARACTERISTICS

Parameter		Value
Biogas Production	m³(STP)/d	2986
Utilised	%	100
Wasted	%	0
Methane Concentration	%	61.8
Carbon Dioxide Concentration	%	38.2

4: PURE OXYGEN SUPPLIED

Supply Rates		Unit
Mean	kg(O_2)/h @ 20°C	-
" "	t(O_2)/d @ 20°C	-
Summer	" " @ 25°C	-
Winter	" " @ 15°C	-

5 COST ANALYSIS

Cost Item (SAR 1,000,000's)		Cost
Full Plant Capital	1995	6.64
Operating (inc. Cap. Repayment)	2002	1.31
Upgrade Capital	1995	3.64
Operating (inc. Cap. Repayment)	2002	0.73

Table 8.4 Predicted Operational Data and Cost Breakdown for Different System Configurations

SYSTEM CONFIGURATION (Treating 200m³/d, T_{amb}=20°C)	Oxygen Injection Rate kgO/h	Diesel Fuel Rate ℓ/h	Process Temperatures °C		Percentage % Volatile Solids Removal			Biogas Production m³/d		Final Sludge Solids Concentrations kg/m³			Pasteurisation	Heating Back-Up	Capital Costs (SAR's x10⁶)		Operating Costs (Inc. Cap.;SAR's x10⁶)	
			Aerobic	Anaerobic	Aerobic	Anaerobic	Overall	Produced	Used	BVS	VS	TS	✓ or ✗	✓ or ✗	Upgrade	Full Plant	Upgrade	Full Plant
A Dual Digestion: No external heating	118	0	63	53	20.6	39.7	52.1	1961	0	0.3	15.5	23.1	✓	✗	0.71	3.71	2.16	2.74
B Dual Digestion: Interstage heat exchange	65	0	60	35	11.3	33.3	40.9	2064	0	3.5	19.1	26.7	✓	✗	0.74	3.74	1.60	2.18
C Dual Digestion: Afterstage heat exchange	76	0	63	53	13.2	43.1	50.7	2326	0	0.4	16.0	23.6	✓	✗	0.73	3.74	1.96	2.54
D Dual Digestion: Supplementary heat with hot water boiler	58	0	63	53	10.1	44.4	50.0	2483	977	0.4	16.2	23.8	✓	✓	1.23	4.23	1.85	2.43
E Dual Digestion: Supplementary heat from gas engine	58	0	63	53	10.1	44.4	50.0	2483	2483	0.4	16.2	23.8	✓	✓	3.24	6.25	1.19	1.77
F Mesophilic Digestion: Heating with hot water boiler	0	0	-	35	-	41.8	41.8	2750	1415	2.5	18.8	26.4	✗	✗	1.42	7.46	1.46	2.62
G Mesophilic Digestion: Using heat from gas engine	0	0	-	35	-	41.8	41.8	2750	2750	2.5	18.8	26.4	✗	✗	2.66	8.67	0.59	1.75
H Thermophilic Digestion: Heating with hot water boiler	0	0	-	53	-	48.0	48.0	2986	2384	0.5	16.8	24.4	✓	✗	1.27	4.28	0.74	1.33
I Thermophilic Digestion: Using heat from gas engine	0	18	-	53	-	48.0	48.0	2986	2986	0.5	16.8	24.4	✓	✗	3.64	6.64	0.73	1.31

8.5.7 Evaluation of the Simulated Performance of Different Sludge Treatment Systems

Comparison between the different system configurations is made in terms of (1) final sludge quality, (2) process stability, and (3) system costs.

Final Sludge Quality

In each case where the anaerobic digester is operated at thermophilic temperatures (**A**, **C**, **D**, **E**, **H**, and **I**) the quality of the final sludge (in terms of remaining BVS concentration) is superior to that of the sludge treated by conventional mesophilic digestion (**F** and **G**). For each dual digester case where the anaerobic stage is at thermophilic temperatures (with R_h= 10d) the percentage volatile solids removal is in excess of 50%.

The only configuration with a final sludge quality inferior to that of conventional mesophilic digestion (R_h=20d) is the dual digestion system with Interstage heat exchange (**B**) with the digester operated at mesophilic temperatures (R_h=10d). To improve the final sludge quality the anaerobic digester retention time would need to be increased from 10 days to about 18 days (an aspect discussed in Section 8.5.2 above). This substantiates the earlier point raised that the anaerobic digester retention time in dual digestion is not so much governed by the stability of the anaerobic processes in the digester but rather the stability of the final sludge product (in terms of residual BVS concentration, ***SOUR*** or %VS removal). Ten days retention time is sufficient for sludge stability with thermophilic digestion but longer retention times (15 to 20 days) are required for mesophilic digestion. The increase in digester capacity with mesophilic digestion would result in higher capital costs.

Except for conventional mesophilic digestion (**F** and **G**), each system is able to pasteurise the sludge during treatment. For the dual digestion systems where the digester is operated at thermophilic temperatures (**A**, **C**, **D**, and **E**), the anaerobic stage provides some back-up should incomplete pasteurisation occur in the aerobic stage (i.e. as a result of a drop in reactor temperature or short circuiting etc.).

System Stability

Each of the dual digestion systems described can be regarded as stable. The principal concern in the operation of the aerobic reactors is the reliability of the recirculation pumps. Consequently, to provide back-up, a stand-by pump has been included in the capital cost and is recommended for practical implementation.

Thermophilic anaerobic processes have in the past often been regarded as less stable than mesophilic processes. However, recent studies by Ahring (1994) have disproved this

belief. Thermophilic anaerobic digesters were shown to be just as stable as mesophilic digesters, although Ahring (1994) recommends that precautions should be applied during start-up. The application of aerobic pre-treatment prior to thermophilic digestion in the dual digestion process would be entirely suitable and beneficial in this regard. The transition to thermophilic temperatures during phase II, proceeded without undue concern. Whilst the volatile acid alkalinity showed a slight increase during this period, there was sufficient buffer capacity in the aerobically pre-treated feed sludge to keep the anaerobic process stable.

Perhaps the principal reason why the stand alone thermophilic digestion process has not become preferable to mesophilic digestion for the treatment of sewage sludge (there is sufficient incentive to implement thermophilic digestion from the significant capital cost savings which can be made as a result of reduced digester capacity requirements) is because of the susceptibility of the thermophilic process upset with temperature fluctuations. Dependence on a single heat source subject to sporadic failure therefore is a likely source of thermophilic process failure; With feed sludge pre-heating, failure of the heating system will result in cold feed sludge passing to the digester resulting in a sharp temperature decrease. If biogas is used to fuel the heating system, the problem is compounded further, as a thermophilic digester upset will cause the biogas production to decrease. In contrast, in the dual digestion system with thermophilic digestion, should a problem arise with the biological heating in the aerobic reactor (e.g. recirculation pump failure, insufficient oxygen etc.), the digester is protected somewhat from receiving a shock of cold feed sludge due to the heat retention of the aerobic reactor. The temperature of the feed sludge to the digester (ex aerobic) would show a gradual decline thereby providing a certain degree of protection against digester failure. Indeed, at Athlone the transition to and from thermophilic temperatures was relatively smooth and took place without affecting the stability of the anaerobic biological processes.

Taking cognizance of the problem of temperature control for the thermophilic anaerobic digestion process, the dual digestion system with supplementary heat provided by a boiler or gas engine (fuelled with biogas) (**D** and **E**) is considered the most suitable as two independent sources of heat are available to maintain the digester temperature as close as possible to 53°C.

System Costs

For the basic dual digestion system (**A**) the capital costs are approximately half that of conventional mesophilic anaerobic digestion (**F**) as a consequence of a reduction in required anaerobic digester capacity. The total process cost (including repayment on the capital) is similar for both processes due to the high oxygen costs.

The economic feasibility evaluation demonstrates that it is cost effective, when operating an anaerobic digester, to derive maximum benefit from the biogas generated by running a gas engine. For conventional mesophilic digestion using a conventional boiler (**F**) whilst capital costs are 16% lower in comparison to the gas engine configuration (**G**), the operating cost is 33% higher.

The most cost-effective configuration for the dual digestion system is that incorporating the gas engine and utilising the waste heat to supplement the biological heating rate (**E**). The process costs of this configuration are 35% lower than the basic dual digestion process (**A**).

Overall, the cheapest configuration to operate (including capital repayment) is that of conventional thermophilic digestion (**H** and **I**). The saving is brought about by (1) reduced digester capacity in comparison with mesophilic digestion, and (2) at an ambient temperature of 20°C, the biogas produced is sufficient to meet all the fuel requirements for heating the system. There is little difference in cost as to whether a boiler or gas engine is used, the reason being that with the gas engine there is insufficient biogas to meet the fuel requirements and additional fuel is required (in this case a dual fuel engine would be required).

Favoured Configuration

In terms of process and product stability and costs, the favoured configuration is the dual digestion system with supplementary heat from the gas engine (**E**). The predicted %VS removal with thermophilic digestion is at 50% (equivalent to that achieved after 30 days mesophilic digestion). The thermophilic anaerobic digestion process is protected somewhat against process upset as a result of a loss of heat input as two independent methods of heating the sludge are applied: (1) biological heating in the aerobic reactor, and (2) recovery of heat from an installed gas engine. In the event of failure of the pure oxygen injection system (i.e. recirculation pump failure, insufficient oxygen) the exhaust heat from the engine can be called upon to provide a large part of the heating requirements (system would switch to configuration **I**). Should the gas engine suffer from mechanical breakdown (or be shutdown for routine maintenance), the system could be switched to configuration **A**, with biological heating in the aerobic reactor providing all the heating requirements for the digester. Also, by making provision at the design stage, heat can be temporarily recovered from the hot effluent sludge (**C**) if the operation of the gas engine is stopped for some reason.

The System **E** evaluation was based on a set 1:1 ratio of biological heating to external heating. The oxygen supply rate (and therefore operating cost) could be reduced further

by increasing the proportion of external heat. A further reduction in operating costs can be achieved for system **E** by the use of a Vacuum Swing Adsorption (VSA) plant (see Appendix 14), which produces oxygen at a cost of approximately 50% that of liquid oxygen. This configuration is particularly suited to VSA oxygen generation because the oxygen supply rate can be kept constant throughout the year, and therefore be provided fully by a VSA plant (liquid oxygen need only be used if operation of the VSA plant is stopped); The additional heating required during the winter months could be provided by the heat from the gas engine. In contrast for configurations **A-C**, because the oxygen requirement fluctuates between the summer and winter periods, the VSA plant would be designed to meet the minimum oxygen requirement. The additional oxygen requirement would be have to be provided by liquid oxygen as there is no alternative heat source to absorb the seasonal variation in the heat requirement.

The comparison of dual digestion system configuration **E** (thermophilic digestion) with conventional mesophilic anaerobic digestion **F** is summarised below, in terms of perceived advantages and disadvantages.

Advantages of System E:
Pasteurised Sludge (double stage).
Higher degree of volatile solids removal (50% vs 42%).
Reduction in capital cost due to the reduction in required digester capacity.
Back-up heating source for thermophilic digestion.
Ideal for VSA plant application.

Disadvantages of System E:
Higher operating cost, although overall cost (incl. capital repayment) is similar.
Reduces biogas production by approximately 10%.
Digester structure must be able to withstand thermophilic temperatures.
Feed Sludge needs to be macerated.

Whilst it is recognised that the determination of system costs quoted in this feasibility evaluation are rough approximations and largely empirical, it demonstrates that when considering the installation or upgrade of an anaerobic treatment plant it is beneficial to consider the various dual digestion configurations. Application of the general dual digestion model allows the cost effectiveness of different system configurations to be evaluated and assists in identifying those configurations for which further in-depth cost appraisal is warranted. A more precise estimate of system costs can then be made by evaluating the individual needs for each specific site and using actual quoted

manufacturers costs for each item. General conclusions from this sludge treatment system performance and economic feasibility evaluation are given in Chapter 9.

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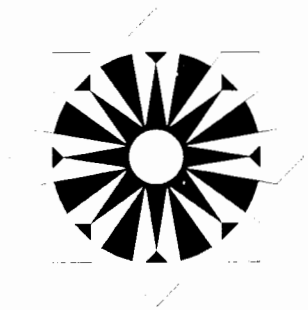

CHAPTER 9

CONCLUSIONS AND RECOMMENDATIONS: PHASE II

9.1 OVERALL ASSESSMENT OF PHASE II

Phase II of the study, to investigate the dual digestion process using a combination of both air and pure oxygen to pasteurise and stabilise sewage sludge, was successful in that all the initial objectives laid out for the project were achieved. The plant operated for a 152 day evaluation period, commencing on the 16th July 1994 (designated day 1 of the evaluation period) and lasting until the 14th December 1994 (day 152). Between days 41 and 73 the aerobic reactor recirculation line was out of commission to allow the pipework to be modified. Outside of this period the plant operated continuosly, without any major mechanical problems. Consequently sufficient data was collected to fully evaluate the system. Unfortunately foaming did not occur inside the reactor, and therefore could not be exploited in the manner envisaged - i.e. to improve the *OTE* of the air oxygenation system.

As in the case of phase I, special emphasis was placed on examining the performance of the aerobic reactor because of the fundamental importance of biological heat generation. A steady state (liquid and gaseous) mass and heat balance over the aerobic reactor was formulated which accounted for pure oxygen and/or air oxygenation. The consistency of the value obtained for the specific heat yield (Y_h) for eleven different steady state periods under widely differing operating conditions indicated that the mass and heat balance equations were sufficiently general and sensitive to the changing operating conditions. Indeed, the close agreement with the Y_h value determined during phase I (when oxygenation was with air alone) and with that obtained by previous workers on a pure oxygen system placed a great deal of confidence in the defined process stoichiometry. The model developed from this can therefore be reliably used for the design and evaluation of aerobic reactors aerated with air and/or pure oxygen.

The central and most important information required for design and evaluation of the aerobic reactor is a knowledge of the oxygen transfer characteristics (such as oxygen transfer rate and efficiency at different air/oxygen supply rates) of the aeration device(s). Operation of the sludge recirculation line under different sets of flow conditions yielded valuable information with regard to defining appropriate flow criteria in order to achieve a satisfactory level of pure oxygen transfer efficiency. Both the air and oxygen oxygenation systems were found to operate independently of each other, which made modelling the biological heat process via the specific heat yield Y_h simple.

Volatile solids destruction (removal) in the aerobic reactor was significant and could be linked to the rate of oxygen utilisation. This relationship formed the basis for predicting VS destruction from the required ***OUR*** for specific aerobic reactor heat balance conditions. Together with a kinetic model developed to simulate the anaerobic process, the aerobic reactor model allowed the fate of the volatile solids to be described through the dual digestion system i.e. effect of VS oxidation (removal) in the reactor on subsequent biogas production in the digester.

During phase II, it was possible to operate the digester at short retention times to test the claim made for the dual digestion process that it was possible to reduce retention time. The high sensible heat content of the sludge from the reactor forced the digester into the thermophilic region as no interstage heat exchange was available. At temperatures between the mesophilic and thermophilic range, the digester performance was adversely affected i.e. reduced VS removal. Nevertheless, evaluation of digester performance was conducted in terms of both process stability and final sludge product stability.

9.2 CONCLUSIONS DRAWN FROM PHASE II

The initial objectives planned for phase II, were described in Chapter 1. The conclusions drawn from phase II are divided into categories in keeping with the laid down objectives. The categories considered are

1 Characteristics of the Aerobic Reactor Oxygenation Systems
2 The Biological Heating Rate and the Steady State Heat Balance
3 Aerobic Reactor Retention Time
4 Volatile Solids Destruction in the Aerobic Reactor
5 Conditioning Effects of Aerobic Pre-Treatment
6 Requirements for Sludge Disinfection
7 Requirements for Heating the Anaerobic Digester
8 Dual Digester Process Stability
9 Final Sludge Stability and Dewaterability
10 Modelling of the Dual Digestion Process
11 Operation of the Dual Digester from a Practical Viewpoint
12 Capital, Operational, and Maintenance Costs
13 The Viability of the Dual Digestion Process using Air+Pure Oxygen

The conclusions drawn in each category are preceded by a brief discussion to place their significance in context with the defined objectives.

9.2.1 Characteristics of the Aerobic Reactor Oxygenation Systems

It was accepted for the purposes of this investigation that the oxygen utilisation rate *OUR* was the most appropriate parameter for monitoring the rate of biological heat generation H_b in the thermophilic aerobic reactor in the dual digestion system. To measure the *OUR*, the characteristics of the two separate oxygenation systems (air and pure oxygen) needed to be well defined. Accurate estimation of the *OUR* was made by conducting a gaseous mass balance across the aerobic reactor. The balance was solved by measuring the oxygen volumetric fraction in the effluent gas stream, accepting the respiration quotient value Y_{CO2} determined during phase I, and making an initial assumption that the two oxygenation devices acted independently. The subsequent accuracy of the steady state heat balance proved this approach to be correct.

Conclusions:

1.1 The two separate oxygenation systems (air and pure oxygen) were found to act independently of each other. Each system contributed independently to the overall oxygen transfer rate (i.e. $OTR=OTR^{AIR}+OTR^{O2}$). Likewise, the oxygen utilisation rate *OUR* could be considered to consist of separate contributions from the oxygen derived from the *air* stream and from *pure oxygen* injection (i.e. $OUR=OUR^{AIR}+OUR^{O2}$). The initial assumption made that the two oxygenation devices act independently was therefore correct.

1.2 The aerobic reactor operated under oxygen limiting conditions during each steady state period (confirmed by the absence of dissolved oxygen in the reactor sludge). Under such conditions, the biological oxygen utilisation rate *OUR* was limited by the combined oxygen transfer rate *OTR* delivered by the two oxygenation devices (i.e. *OUR=OTR*).

1.3 During phase II no foaming occured in the aerobic reactor, consequently the oxygen transfer efficiency of the air oxygenation device remained relatively constant in the range OTE^{AIR}=**11.9±0.4%**. During phase I, under non-foaming conditions with one compressor operating, values for OTE^{AIR} were in the range 12.6±0.3%. The reduction in the mean OTE^{AIR} from 12.6% to 11.9% between phase I and phase II, is attributed to a deterioration in the condition of the course bubble diffusers.

1.4 The maximum dry air flow rate which the air liquid ring compressor could deliver during phase II was **680** m^3(STP)/h. This corresponds to a volume specific oxygen

supply rate OSR^{AIR} of **1.025** kg(O_2)/m^3.h. At this supply rate the average oxygen transfer efficiency OTE^{AIR} was **11.8**% giving rise to an oxygen utilisation rate OUR^{AIR} = **0.121** kg(O_2)/m^3.h.

1.5 The observed oxygen transfer efficiency of the pure oxygen oxygenation system OTE^{O2} was strongly dependent on the flow conditions in the sludge recirculation line. The flow characteristics necessary to achieve an $OTE^{AIR+O2} > 80\%$ were found to be:

- Minimum sludge flow rate $Q_{min} \geq$ **690** m^3/h
- Pressure head at injection point $h_{min} \geq$ **23.0** m
- Velocity at point of discharge $v_{min} \geq$ **13.2** m/s

1.6 The maximum oxygen transfer rate OTR^{O2}_{max} effected by the pure oxygen system during the evaluation period was **0.435** kg(O_2)/m^3.h, which at an transfer efficiency OTE^{O2} of **83.3**% corresponds to an oxygen supply rate OSR^{O2} of **0.537** kg(O_2)/m^3.h. The system was not tested to determine if significantly higher OTR^{O2} rates than 0.435 kg(O_2)/m^3.h could have been achieved as this would have resulted in the reactor overheating (>65°C). At 0.435 kg(O_2)/m^3.h, the OTR^{O2}_{max} was about 10% higher than that which is theoretically calculated from the solubility of pure oxygen in pure water under the conditions (temperature, pressure and flow rate) at the injection point. It would appear that either some supersaturation occurs or the utilisation of oxygen is such that redissolution of precipitated oxygen takes place before it escapes with the vent gas.

1.7 The maximum biological oxygen utilisation rate *OUR* recorded during the evaluation period was **0.498** m^3/kg(O_2).h (where OUR^{AIR}=0.063 and OUR^{O2}=0.435). At this time the average total solids concentration of the primary sludge was 42kg(TS)/m^3 and the retention time in the aerobic reactor was 0.96 days. The fact that the reactor remained oxygen limited at this high *OUR* demonstrates that the aerobic process could possibly support higher *OUR*'s (up to the point where OUR^{max} is reached). Estimates of OUR^{max} (when the reactor becomes substrate limited) were made from the developed kinetic model which describe the rates of enzyme hydrolysis and readily biodegradable volatile solids generation for utilisation by the thermophilic biomass.

1.8 The value for the respiration quotient Y_{CO2} (defined as moles of carbon dioxide generated per mole of oxygen utilised) measured during phase I was accepted for use during phase II to solve the gaseous mass balance. The close correspondence

of the specific heat yield Y_h value from the steady state heat balance for phase II, provided indirect evidence that the Y_{CO2} value for phase II also closely corresponded to that of phase I.

9.2.2 The Biological Heating Rate and the Steady State Heat Balance

The biological heating rate H_b is of fundamental importance to the performance of the aerobic reactor. Calculation of H_b by difference from the steady state heat balance allows the specific heat yield Y_h to be determined. Once Y_h is known the heat balance can be used to predict aerobic reactor temperatures under a wide variety of operating conditions.

Conclusions:

2.1 The biological heating rate H_b was directly proportional to the oxygen transfer rate ***OTR*** which, under oxygen limiting conditions, fixes the biological oxygen utilisation rate ***OUR***. The constant of proportionality is the specific heat yield Y_h.

2.2 The specific heat yield Y_h (defined as the quantity of biological heat produced per mass of oxygen utilised) was measured to be **12.8** MJ/kg(O_2) during phase II. This is the same value as that determined during phase I and shows good agreement with the value of 13.0 obtained by Messenger *et al* (1992). The spread in Y_h values between each steady state period was small (12.3 to 13.5 MJ/kg(O_2)) in spite of the widely differing operating conditions during the 11 steady state periods.

9.2.3 Aerobic Reactor Retention Time

It was anticipated that the use of a pure oxygen injection system, to supplement the air oxygenation system, would provide improved reactor temperature control and allow operation of the aerobic reactor at a 1 day retention time instead of 4-6 days with air only.

Conclusions:

3.1 With the aerobic reactor operating under oxygen limiting conditions, the reactor sludge temperature could be completely and instantaneously controlled by the pure oxygen supply rate OSR^{O2}.

3.2 With the reactor temperature controlled at >60°C, the shortest retention time at which the reactor was operated was **0.96** days.

3.3 The estimated minimum retention times (R_{ft}^{min}) for operation before the reactor becomes substrate limited for different sludge types are as follows:

- Primary sludge **0.4** days†
- Primary sludge/humus sludge mixture **1.1** days
- Primary sludge/waste activated sludge mixture **1.2** days
- Waste activated sludge **2.2** days

† biomass growth rate limitation will occur before this retention time is achieved (see Item 3.4 below)

Estimate of R_{ft}^{min} is based on calculated RBVS generation rates obtained from the developed VS removal kinetic model and is principally dependent on sludge type, the VS loading rate, and the required ***OUR*** to maintain a certain set of conditions. The above estimates are based on an ambient (feed sludge and surrounding air) temperature of 20°C, feed sludge total solids concentration of 40kg/m^3 and aerobic reactor temperature +60°C.

3.4 It is estimated that the minimum retention time at which it is possible to operate the reactor before the thermophilic aerobic bacteria start to be washed out from the reactor is around **0.8** days due to maximum specific growth rate limitation. This will cause a reduction in ***OUR*** for reasons other than substrate or oxygen limitation. In order to incorporate a factor of safety a 1 day retention time is regarded as the practical minimum.

9.2.4 Volatile Solids Destruction in the Aerobic Reactor

A significant degree of volatile solids reduction took place in the aerobic reactor during phase I due to the relatively long retention times (4 to 6 days). As a result it was possible to quantify the VS reduction in terms of the biological heating rate and the oxygen utilisation rate. For the pure oxygen aerobic reactor at Milnerton operated at short (1 to 3 days) retention times, very little VS removal was observed (<1.5%) (Messenger *et al*, 1992). Operation of the aerobic reactor at retention times similar to that at Milnerton during phase II, allowed for further examination of the relationship between VS removal and oxygen utilisation.

Conclusions:

4.1 In contrast to the results obtained at Milnerton, a significant degree of volatile solids destruction took place in the aerobic reactor at Athlone; **20.2%** at an average retention time of 1.6 days i.e. 7.7 kg(VS)/m^3 from an average influent concentration of 38.1kg(VS)/m^3 and flow of 115m^3/d.

4.2 The average quantity of biological heat produced per mass of volatile solids destroyed was **19** MJ/kg(VS). This shows relatively good agreement with the value of 22 MJ/kg(VS) determined during phase I and the value of 21 MJ/kg(VS) obtained by Andrews and Kambhu (1971), which has become a recognised standard value for the design of the ATAD process.

4.3 The average quantity of oxygen consumed per mass of volatile solids destroyed was calculated at **1.6** kg(O_2)/kg(VS). This figure is higher than that normally associated with sewage sludge (1.42 mgCOD/mgVS) but is in close agreement with the ratio of 1.7 kg(O_2)/kg(VS) determined during phase I and the COD/VS ratio of the influent feed sludge which was also 1.7 kg(O_2)/kg(VS).

4.4 The relationship between the rate of oxygen utilisation and the rate of volatile solids destruction was maintained over the full range of operating conditions encountered at Athlone. From a retention time of 1.1 to 4.4 days (oxygenation with air and/or oxygen).

4.4 The average biological oxygen utilisation (or transfer, air + pure oxygen) rates $M(O_2)_{ut}$ determined before and after the modification of the recirculation line during phase II were 34.3 and 75.7 kg(O_2)/h respectively. Although a COD balance requires that the COD removal and oxygen utilisation rates should be equal, the $M(O_2)_{ut}$ rates were **45%** and **36%** lower respectively than the average COD removal

rates for the same periods (62.5 and 118.1 kg(COD)/h). For comparison, during phase I, $M(O_2)_{ut}$ was approximately **12% lower** and at Milnerton **234% higher** than the COD destruction rate.

4.5 For operation of the Athlone aerobic reactor, whether with air or a combination of air and pure oxygen, at retention times ranging from 1 to 6 days, a carbon balance of **80-89%** was achieved i.e. of the carbon VS (assumed to be $C_5H_7O_2N$) entering the system via the sludge liquid stream, 80-89% was measured in the effluent gas and liquid streams. Details of the carbon balance are given in Appendix 6.

4.6 The consistent relationship between the oxygen utilisation, VS removal, and heat generation rates permitted an integrated mathematical model to be developed linking these rates to the retention time and reactor temperature. The predictive power of the model is discussed in Section 9.2.10 below.

4.7 With the aid of the integrated model, the increase in ammonia concentration could be estimated from the %VS removal (see Section 9.2.5 below). Also, the effect of VS reduction in the aerobic reactor on subsequent biogas production in the anaerobic digester can be predicted.

9.2.5 Conditioning Effects of Aerobic Pre-Treatment

As higher aerobic reactor loading rates were possible during phase II, the retention time in the anaerobic digester could be reduced to below 10 days. This provided the opportunity to assess the conditioning effects of aerobic pre-treatment and their impact on the stability of the anaerobic digestion process and quality of the final sludge product.

Conclusions:

5.1 As a result of aerobic treatment the average ammonium and bicarbonate alkalinity concentrations increased from 108 mgN/l and 180 mg$CaCO_3$/l in the feed to 518 mgN/l and 850 mg$CaCO_3$/l in the reactor respectively. The pH increased from an average 5.4 to 7.2.

5.2 Approximately 9% of the carbon dioxide generated through the mineralisation of organics remained in solution and reacted with ammonium ions to form ammonium bicarbonate which increased the alkalinity.

5.3 No nitrification took place in the aerobic reactor; had this taken place the alkalinity would have been reduced to virtually zero because the alkalinity generation arises principally from the ammonification of proteins ($NH_3 \rightarrow NH_4$).

5.4 Both the ammonium ion concentration and the bicarbonate alkalinity produced in the aerobic reactor were approximately proportional to the concentration of volatile solids destroyed for the range of operating conditions.

The stability of both the anaerobic digestion process and the final sludge product after treatment at the short retention time is discussed in Sections 9.2.8 and 9.2.9 below.

9.2.6 Requirements for Sludge Disinfection

To achieve disinfection the feed sludge has to be exposed to sufficiently high temperatures for specified periods of time without short circuiting. It was important therefore to evaluate the autoheating potential of the aerobic reactor in treating primary sludge at short retention times. With the digester operating in the thermophilic region the disinfecting value of the digester was evaluated.

Conclusions:

6.1 Sufficient autoheating potential exists for thermophilic temperatures (>60°C) to be maintained in the aerobic reactor throughout the year at an operating retention time of 1 day.

6.2 Disinfection in terms of a reduction in faecal coliform concentration to $<10^3$/100ml was achieved in the reactor when the temperature reached 58°C at a 2 day retention time (flow rate 100m^3/d).

6.3 When the sludge flow rate was increased to 192m^3/d (1 day retention time) the faecal coliform concentration in the reactor sludge increased to between 10^4 and 10^6/100ml. This suggesting that at the higher sludge loading rate, short circuiting was taking place in the aerobic reactor. **To provide effective pasteurisation feeding on a draw and fill basis would be required**, rather than a displacement flow through batch basis as was done at Athlone.

6.4 The faecal coliform concentration dropped to $<10^3$/100ml in the digester sludge when the digester temperature entered the thermophilic range at a flow rate of 100m^3/d. It should be noted that at this time the faecal coliform concentration in

the reactor sludge was low ($<10^3$/100ml; see item 6.2 above) which could account for the low levels.

6.5 The concentration of faecal coliforms in the digester sludge was in excess of 10^3/100ml at the high sludge loading rate (196m^3/d, 9 day digester retention time) when higher concentrations of faecal coliforms (10^4-10^n/100ml) were discharged from the aerobic reactor (see item 6.3 above).

6.6 An average of 4 orders of magnitude reduction in faecal coliforms was observed across the aerobic reactor (from 1.8 x 10^9 to 1.1 x 10^5/100ml at 3.8% and 3.0% volatile solids concentration respectively) with a further 2 orders of magnitude reduction after anaerobic digestion (to 1.2 x 10^3/100ml at 1.8% volatile solids concentration).

6.7 Configuring the dual digestion system so that both the aerobic and anaerobic stage operate at thermophilic temperatures would prove beneficial in guaranteeing long term pasteurisation. The provision of two stages of pasteurisation, each capable of achieving complete pasteurisation independently, would provide the necessary cover in the event of any adverse deviation from normal operating practice (i.e. through accidental short circuiting, or a temporary drop in reactor temperature due to a shortage of oxygen).

9.2.7 Requirements for Heating the Anaerobic Digester

The operating temperature of the anaerobic digester in the dual digestion system will be dependent upon (a) the sensible heat content of the sludge entering the digester, governed by the flow rate and temperature of the sludge, and (b) the ambient temperature. Operating conditions in the aerobic reactor therefore have a definite bearing on the subsequent operating temperature of the digester.

Conclusions:

7.1 With the high ambient temperatures encountered in South Africa, to operate the digester at a retention time of 10-15 days in the mesophilic region, interstage cooling would be required.

7.2 Alternately, the digester could be operated in the thermophilic region at a retention time of 10 to 15 days by controlling the aerobic reactor temperature at a sufficiently high value (>63°C). With a simulation model compiled for the dual digestion system (see 9.2.10 below) it is possible to estimate the required operating

temperature of the reactor to fix the digester temperature at 53°C provided the heat balances for the aerobic reactor and anaerobic digester are well defined..

7.3 If no allowance is made of the interdependence of the operating temperatures of the reactor and digester then there is a high risk that the anaerobic digester temperature will, at the ambient (sludge feed and surrounding air) temperatures (15°-25°C) in South Africa, fall between the mesophilic and thermophilic temperature regions.

9.2.8 Dual Digester System Stability

Aerobic pre-treatment is claimed to reduce the minimum required retention time for operation of the anaerobic digestion stage; retention times as low as 8-10 days are claimed to be possible. It is important therefore to examine the stability of the process at the short retention times encountered during phase II (20-9 days).

Conclusions:

8.1 The shortest retention time at which the digester operated was 9 days, with the digester temperature in the *thermophilic* region at 53-55°C. The stability of the anaerobic processes during this period was considered good after acclimatization to thermophilic temperatures (see item 8.2 below), with the pH~7.3, HCO_3^- alkalinity ~2700mg($CaCO_3$)/l and volatile acid alkalinity~400mg($CaCO_3$)/l).

8.2 Two events, (1) the transition between mesophilic and thermophilic temperatures and (2) the reduction in retention time from 18 to 9 days, caused the volatile acid alkalinity to temporarily increase to around 1000mg($CaCO_3$)/l with a concommitent decrease in pH to around 6.9. Sufficient buffer capacity was however available to absorb the effects.

8.3 Thermophilic anaerobic digesters have in the past, been regarded as less stable than mesophilic processes. However, recent studies by Ahring (1994) have disproved this belief; if precautions are taken during start-up and in maintaining satisfactory control over the digester temperature, thermophilic operation can be as stable as mesophilic operation.

8.4 By operating the aerobic reactor at a minimum temperature of 60°C (to achieve pasteurisation), under South African conditions (ambient temperature 15° to 25°C) without interstage cooling the anaerobic digester (at retention times <20days) temperature will be in the region 50° - 55°C. Control of the aerobic reactor

temperature at >60°C with the pure oxygen supply rate will allow the anaerobic digester temperature to be controlled at say 53°C, the recommended temperature for thermophilic operation (Ahring, 1994).[1]

8.5 When operating the anaerobic digestion stage at thermophilic temperatures, it is recommended that external heating of the aerobic reactor be incorporated (using biogas as fuel). This is not only cost effective in terms of reducing the oxygen supply rate and making beneficial use of the biogas generated (via a gas engine), but also provides two independent sources of heat for the reactor, each being a back up for the other. The dependence on a single source of heating is perhaps the principal reason why thermophilic digestion has become regarded as unstable and as a result not achieved widespread usage.

8.6 One of the claimed benefits of the dual digestion system is that aerobic reactor heat pre-treatment of the sludge allows the anaerobic digester to operate at short retention times (~10 days). This is considered true only if the digester temperature is in the thermophilic range, in which case a sufficiently stable sludge is produced; at mesophilic temperatures, a retention time of 15 days or longer is required to produce a sludge of equivalent stability to that from conventional mesophilic digestion. Consequently it is not the stability of the anaerobic process per se that governs the minimum retention time but the quality required for the final sludge product.

8.7 The aerobic reactor is an appropriate pre-treatment stage for the thermophilic digester because it provides the necessary temperature and pH buffering to allow stable operation in the thermophilic range.

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[1] Ahring (1994) found that the optimal temperature for thermophilic anaerobic digestion is 60°C. However, for practical operation a temperature of between 52° and 56°C is recommended.

9.2.9 Final Sludge Stability and Dewaterability

Other than disinfection, the two other factors which are of paramount importance when evaluating a sludge treatment system are (1) the final sludge stability and (2) the dewaterability. Both apects were evaluated during phase II. The quality of the final sludge was examined in terms of stability (%VS removal and Specific Oxygen Utilisation Rate *SOUR*), fermentability (gas production), odour and dewaterability (Specific Resistance to Filtration, SRF).

Conclusions:

9.1 The final sludge from the digester was stable and did not undergo further fermentation. The volatile solids content of the final sludge was **67**% (dry mass basis). The average percentage TS, VS and COD removals in the aerobic reactor were 16, 20 and 28% respectively and in the anaerobic digester were 30, 40 and 37% respectively. This was at an overall average of 1.6 days aerobic reactor retention time and 16 days anaerobic digester retention time.

9.2 The average percentage TS, VS and COD removals in the system overall were 42, **52** and 55% respectively and are comparable to conventional anaerobic digestion.

9.3 A stable sludge is regarded as one on which a %VS removal in excess of 38% has been achieved, and has a residual ***SOUR*** of <1g(O)/kg(TSS).h. With respect to the %VS removal the sludge would be regarded as stable. An ***SOUR*** test was performed on the final sludge from the process after the evaluation period ended. At the time of the test, the aerobic reactor retention time was 1 day and the anaerobic digester retention time was 9 days. During the first 24 hours of the test, the ***SOUR*** peaked at 13 g(O)/kg(TSS).h with the average ***SOUR*** being 3.4 g(O)/kg(TSS).h. After 80 hrs the ***SOUR*** had dropped to 1.0 g(O)/kg(TSS).h. Whilst the sludge failed to meet the ***SOUR*** criteria during the early stage of the test, the results were better than that observed for conventional (mesophilic) anaerobically digested sludge. Therefore the sludge stability in terms of ***SOUR*** from the dual digester with a 9 day retention time thermophilic digester was better than that from a 20 day conventional anaerobic digester.

9.4 The final sludge from the digester had an earthy odour which was identical to that from conventional anaerobic digestion. The sludge from the aerobic reactor (as far as could be subjectively established from operating staff) also did not have an offensive odour.

9.5 The dewaterability of the final sludge (SRF = 802 x 10^{12}m/kg) was poorer compared with that of the conventional anaerobically digested sludge at Athlone (218 x10^{12}m/kg) and the dual digested sludge after phase I (368x10^{12}m/kg). However, it was not much worse compared with that from the Milnerton pure oxygen dual digestion system with mesophilic digestion (507 x 10^{12}m/kg) (Messenger *et al*, 1992). It would appear from this that exposure to high temperatures during treatment has a detrimental effect on the dewaterability of treated sewage sludges (as measured by the SRF test).

9.6 During secondary digestion, it was possible to draw off similar quantities of supernatant (approx 50% of the flow) as with conventional anaerobically digested sludge. This is indicative of a significant degree of stability because supernatant withdrawal is not possible if significant digestion continues to takes place in the secondary digester.

9.7 After secondary digestion, the dewaterability of the dual digested sludge reduced to a level (SRF=237 x10^{12}m/kg) similar to that for conventional anaerobically digested sludge at Athlone (218 x10^{12}m/kg) indicating that a period of consolidation improves the dewaterability.

9.2.10 Modelling of the Dual Digestion Process

One of the primary objectives of phase II was to produce a general computer model which could simulate all the main dual digestion system parameters. In addition to being able to define the oxygenation and heating requirements to achieve autoheating using air and/or oxygen, accurate estimates of (1) operating temperatures of both the aerobic and anaerobic stages, (2) volatile solids destruction in both the aerobic and anaerobic stages, and (3) biogas production in the digester, were required.

Conclusions:

10.1 The general model described in Chapter 8 is considered to have successfully met all the important requirements for dual digestion system simulation. Sufficient information was obtained during phases I and II for the simulation model to provide a reasonably reliable description of all the main operating parameters in the system, both for the aerobic reactor (with air or pure oxygen) and the anaerobic digester (at mesophilic or thermophilic temperatures) under a variety of different

system configurations, including interstage and afterstage heat exchange and gas engine or boiler external heat sources.

10.2 The dual digester system model has been compiled in such a way that it can be applied for any size and type of dual digester application, i.e. it is not site specific for Athlone. The principal parameters which are computed by the model are:

- Oxygenation characteristics of the aerobic reactor
- Steady state heat balance for the aerobic reactor
- Heating requirements for the anaerobic digester
- Effect of installing heat exchangers
- Volatile solids destruction in both the reactor and digester
- Biogas production by the digester
- Performance of an installed gas engine or hot water boiler
- The stability of the final sludge
- Minimum reactor retention time to prevent substrate limitation
- Minimum digester retention time to ensure sludge stability

10.3 The model includes a general cost analysis for both operating and capital costs. This allows approximate cost comparisons to be made between different sludge treatment system configurations, such as between the different dual digestion system configurations and operational conditions and conventional mesophilic anaerobic treatment both in terms of the parameters listed above (where applicable) and in terms of approximate capital and operating costs.

9.2.11 Operation of the Dual Digester from a Practical Viewpoint

An important aspect of any system evaluation is to make an estimate of the practical operational problems of the full scale system over an extended period and to make an assessment of its reliability.

11.1 The process proved to be simple to operate. The number of staff required to operate the dual digestion plant was no different from that required for conventional anaerobic digestion. It is recommended that with the digester operated at thermophilic temperatures more frequent monitoring of the digester be carried out.

11.2 The aerobic reactor was easily started. From a starting temperature of 20°C the reactor reached thermophilic temperatures within 8 days. The approximate rate of temperature increase was 4°C per day.

11.3 It is important for the operating staff to ensure that the feed sludge to the reactor is as concentrated as possible to prevent substrate limitation taking place.

11.4 Foaming did not occur during phase II. This absence of foam may be as a result of, (1) pure oxygen oxygenation (as found at Milnerton), (2) reduced sludge recirculation flow rate (700 instead of 1100 m^3/h) or, (3) reduced air flow rate (~500 instead of 760 $m^3(STP)/h$).

11.5 No major mechanical problems were encountered during the evaluation. The pumping rate through the aerobic reactor sludge recirculation line was inadequate during the initial period of phase II due to inadequate sizing and speed of the impellars. Specifications for the pump(s) in order to achieve the correct hydraulic conditions in the recirculation line for >80% ***OTE*** were established. Inspection of the impellars at the end of stage II showed that they were badly worn. This was attributed to the fact that the impellars were constructed from cast iron and were not provided with a protective coating. Ceramic coated shafts and impellars have subsequently been fitted to the pumps, and subsequent wear and tear will be examined in future after a prolonged period of operation.

11.6 In the design and operation of future dual digestion plants it is recommended that a stand-by pump be fitted to the sludge recirculation line to provide back-up in the event of mechanical failure to the first (and to allow routine maintenance to be carried out).

11.7 The course bubble air oxygenation system, although ineffective in contributing to the overall oxygen transfer rate, did contribute the useful function of keeping the reactor aerated on those occassions that pure oxygen injection was interrupted. In this manner the thermophilic aerobic biomass maintained viability. With the re-introduction of pure oxygen injection the recovery time was instantaneous.

11.8 A different more efficient air oxygenation system is recommended, e.g. injection of air into the recirculation stream, with which Wolinski (1985) claims he achieved 100% oxygen transfer efficiency.

11.9 If the feed sludge is to be pre-heated then a macerator will be required to prevent the heat exchanger system from blocking.

9.2.12 Capital, Operational, and Maintenance Costs

An estimation of the capital, operation and maintenance costs for the system was performed to enable comparisons of different dual digestion system configurations to be made with conventional anaerobic digestion.

One of the principal objectives in constructing the general dual digestion simulation model was to be able to make an assessment of capital and operating costs.

Conclusion:

12.1 The cost analysis performed by the general dual digestion simulation model is considered adequate to enable an initial cost assessment of a variety of dual digestion systems to be made.

12.2 In a cost assessment of a variety of dual digestion configurations for application under South African conditions (average ambient temperature ~20°C) the favoured configuration incorporates supplementary heating of the aerobic reactor from the waste heat from a gas engine (which is used to generate electrical energy). Using pure oxygen for oxygenation, the aerobic reactor retention is at 1 day with the anaerobic digester retention time at 10 days. The overall cost of the process compares favourably with conventional mesophilic digestion.

12.3 For any institution considering installing, or enlarging, a conventional anaerobic digestion plant, it is recommended that a cost analysis be performed to determine the cost-effectiveness of the dual digestion process. It may be that dual digestion can produce a better quality product at a cheaper overall cost.

12.4 The upgrading of a conventional anaerobic digestion plant to dual digestion is a relatively simple task involving the addition of an appropriately sized aerobic reactor with the necessary set of mechanical equipment to drive the process. The principal capital cost involved in the upgrade therefore will be based primarily on the required size of the aerobic reactor (+ mixing pumps; high power is required) This can be determined from an application of the steady state heat balance (see Chapter 8) and will be dependent upon:

- The required reactor operating temperature.
- The range of sludge feed and ambient temperatures.
- The volume of sludge to be treated.
- Whether pure oxygen, air, or oxygen enriched air is used for aeration.
- The process volume of the existing anaerobic digester.
- The ***OTR-OTE*** characteristics of the oxygenation device.

With the digester operated at thermophilic temperatures, the use of a conventional boiler system or gas engine (fuelled with biogas) is recommended to provide a heating back-up (this is the optimum recommended configuration, see item 12.2). This will increase the capital outlay. In upgrading from convential digestion, it may well be that adequate biogas heating facilities will already exist. Whilst increasing the capital cost, the use of supplementary heat will reduce oxygen requirements and therefore lead to lower operating costs.

9.2.13 The Viability of the Dual Digestion Process

The fundamental factors in deciding whether or not a particular treatment system is viable for wide application are (1) is it capable of producing a final sludge product that conforms to the criteria laid down for disposal? (2) the ease of operation and its reliability, (3) the cost factor (how competitive is it, with regard to capital, operational and maintenance costs, in comparison with other available systems?), and (4) can the system be readily integrated into the existing treatment plant infrastructure?

Conclusions

13.1 **Pasteurisation:** The dual digestion process is capable of producing a pasteurised final sludge product which could be classified as a type D sludge (provided the potentially toxic metal and element concentrations are below the specified limits) in terms of the guidelines for the use of sewage sludge (DNH&PD, 1991).

13.2 **Final Sludge Stability:** If the digester is operated at thermophilic temperatures (53°C) a sufficiently stable sludge is produced after a 10 day anaerobic retention time. At mesophilic temperatures, a retention time of 15 days or longer is required to produce a similarly stable sludge.

13.3 **Dewaterability:** Dewatering the final sludge could be difficult. On the sludge drying beds it dries at about the same rate as well stabilised anaerobically digested sludge. As is the case at Athlone, most dual digestion plants employ thickening and storage tanks following the anaerobic digester stage. Significant quantities of supernatant can be withdrawn providing a fair degree of thickening. In a number of cases, the sludge is used in this thickened liquid form for agricultural purposes (Messenger *et al*, 1992).

13.4 **Ease of Operation and Reliability:** The system proved easy to operate and is considered to be reliable. A stand-by pump is recommended on the aerobic reactor sludge recirculation line to provide back-up in the event of planned maintenance

or breakdown. Further, a seperate air oxygenation system should be provided to keep the reactor contents aerobic in the event of failure to the 'pure oxygen' oxygenation device. If the digester is operated in the thermophilic temperature range then the recommended system configuration is (as described in item 12.2) to employ biological heating and external heating via a gas engine or conventional boiler using biogas as fuel. In this manner two independent sources of heat are available, each being a back-up for the other.

13.5 **System Costs:** For the basic dual digestion system (with thermophilic digester) using pure oxygen without any form of external heating, the capital costs are approximately half that of conventional anaerobic digestion as a consequence of a reduction in the required digestion capacity. The total process cost (including repayment on the capital) is similar for both processes due to the high oxygen costs. By incorporating a gas engine and utilising the waste heat to supplement the biological heating rate (the favoured configuration) total process costs can be reduced by as much as 35%. Further cost savings can be made by generating oxygen on-site using a Vacuum Swing Adsorption (VSA) plant, with oxygen costs reduced by up to 50%. Further, it is important that the oxygenation device employed must be capable of a high oxygen transfer efficiency to ensure that pure oxygen is not wasted. For smaller sized treatment plants, it may be that the basic dual digestion system is favoured, as it may not be economically viable to make use of the biogas which is generated. In any case, each dual digestion system configuration should be evaluated to determine the most cost-effective.

13.6 **System Application:** The upgrading of a conventional anaerobic digestion plant to dual digestion is considered to be relatively simple task (see item 12.4). The design of the plant, in terms of the sizing the aerobic reactor and the required mechanical equipment, can be be readily determined from an application of the simulation model described in Chapter 8. The characteristics of the feed sludge needs to be well defined; Providing the feed sludge is of adequate concentration (>3-4%TS) and contains sufficient energy in the form of readily biodegradable organics (i.e. contains a significant proportion of primary sludge) then aerobic reactor retention times as short as 1 day are possible using pure oxygen. Sludge pre-thickening greatly benefits the process in terms of kg sludge treated as oxygen requirements are based on sludge volume treated.

9.3 RECOMMENDATIONS

9.3.1 Future Application of the Dual Digestion System

Where application of conventional anaerobic digestion is contemplated, whether for new installations or for upgrading existing plants, the dual digestion system should be seriously considered as a possible option. It competes favourably both technically and economically with conventional mesophilic digestion and produces a superior sludge product which can be beneficially utilised in agriculture.

The technical and economical evaluation between the different dual digestion system configurations and conventional mesophilic and thermophilic anaerobic digestion systems can be reliably made by applying the general simulation model for dual digestion systems developed in this report.

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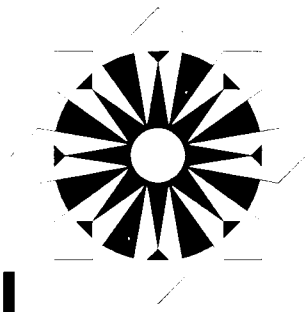

APPENDIX 1

BRIEF HISTORY OF THE DEVELOPMENT OF THE DUAL DIGESTION PROCESS

1a.1 Autothermal Thermophilic Aerobic Digestion

The exploitation of biological autoheating has long been observed in the composting of solid organic wastes. Only recently has this phenomenon been applied to liquid organic slurries. In the early 1970's autothermal thermophilic aerobic digestion (ATAD) emerged as a potentially viable process for the stabilisation and pasteurisation of sewage sludges. Earlier feasibility studies by Andrews and Kambhu (1971) had predicted that sustained thermophilic temperatures were possible if the biological heat were contained and heat losses minimised. Initial success was confined to the use of pure oxygen as vent gas vapour heat losses from the system are minimal. With the introduction of high efficiency aeration devices it became apparent that successful operation using air alone was possible and that the original doubts concerning cooling effects of large volumes of air were unfounded (Jewell and Kabrick, 1980).

With a higher rate of metabolism occurring at thermophilic temperatures relatively short retention times were possible in comparison with conventional processes. Effective pasteurisation occurred within hours, while the stabilisation of the biodegradable sludge solids required process residence times of 5-8 days. Although very short residence times were achieved using pure oxygen, the high oxygen cost factor for stabilisation proved prohibitive and more recent studies on the ATAD process have centred on using air with possible oxygen supplementation to cut down on the oxygen costs (Booth and Tramontini, 1984).

Using pure oxygen, the ATAD process is capable of reaching disinfection temperatures at retention times of the order of 24 hours (Trim, 1984). While little stabilisation takes place under such conditions, the fact that disinfection takes place made it appropriate to consider short residence thermophilic aerobic digestion as a pre-treatment pasteurisation stage to conventional anaerobic digestion. This process became known as Dual Digestion.

1a.2 Dual Digestion Using Oxygen

The first pilot-plant studies into dual digestion using oxygen were undertaken in 1975 by the Union Carbide Corporation. Thermophilic temperatures were achieved in the aerobic reactor at a retention time of 1 day, with satisfactory anaerobic stabilisation occurring within 8 days. The first full scale plant was commissioned in 1980 in the USA and involved the conversion of an existing anaerobic digester (Farrel, 1984). To date the use of the process has been limited to those countries where sludge pasteurisation is obligatory.

A full-scale evaluation of the dual digestion system using pure oxygen was conducted at the Milnerton Wastewater Treatment Plant. The Milnerton research dual digestion plant proved the practicability of the dual digestion process using oxygen under South African conditions. Many of the advantages claimed for the system were verified by the Milnerton research and valuable information was gained in respect of process stoichiometry and biokinetics of heat generation for the aerobic reactor, stabilisation and gas production in the anaerobic digester, sludge quality and dewaterability. (de Villiers *et al.* 1992, Messenger *et al.* 1992). The high cost of using pure oxygen for oxygenation in the aerobic reactor was, however, a disadvantage of the dual digestion process using pure oxygen (Laubscher *et al.* 1992) and it became logical to consider dual digester performance using air.

1a.3 Dual Digestion Using Air

In 1989 the Cape Town City Council initiated a full scale research project to investigate the dual digestion process using air at the Athlone Wastewater Treatment Plant (184 m^3 aerobic reactor and 1800 m^3 anaerobic digester). The general aim of the research programme was to demonstrate the practicability of the dual digestion process, employing air rather than pure oxygen to successfully disinfect and stabilise sewage sludge. The result of that investigation is contained in this report.

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APPENDIX 2

OPERATIONAL, CHEMICAL AND BACTERIOLOGICAL DATA: PHASE I

2a.1 The Evaluation Period: Phase I

Appendix 2 contains a complete tabulation of all the operational, chemical and bacteriological data collected during phase I of the investigation into the dual digestion process at the Athlone Wastewater Treatment Plant. Phase I of the investigation dealt specifically with operation of the aerobic reactor using air alone for oxygenation. Monitoring commenced on the 5th October 1989 (designated day 1 of phase I) and lasted for 45 weeks until the 12th August 1990 (day 312). The operational, chemical and bacteriological data collected during phase II is presented in Appendix 5.

2a.2 System Sludge Stream Monitoring

The feed sludge to the aerobic reactor was drawn from the gravity thickener. An in-line flow meter, placed close to the point of entry to the aerobic reactor, measured the quantity of sludge pumped into the reactor. To take account of the variation of solids concentration in the gravity thickened sludge stream, the plant operators collected a grab sample midway through each feeding session and composited these over 24 hours to make a daily composite sample.

Sludge from the aerobic reactor and anaerobic digester were sampled once every eight hours and composited over 24 hours. The aerobic reactor sludge was sampled at the recirculation mixing pump. The anaerobic digester sludge was sampled at the digester effluent transfer box when sludge was being transferred to the secondary digester.

Samples for bacteriological analyses and dewaterability studies were taken as grab samples on the specific days these analyses were performed.

2a.3 Aerobic Reactor Gas Stream Monitoring

The volumetric influent dry air flow rate to the aerobic reactor $Q(AIR)_{in}$ was monitored by an orifice plate flow meter with the flow recorded on a chart recorder housed in a control room. The flow rate was automatically corrected to STP, 20°C and 760 mmHg as dry air.

The carbon dioxide and oxygen fractions in the aerobic reactor effluent gas stream were monitored approximately three times weekly during each steady state period using a combustion analyser (which also gave the effluent gas temperature).

2a.4 Temperature Measurement

Eight PT100 resistance thermometers were positioned around the aerobic reactor to measure temperatures.

- 3 within the aerobic reactor
- 2 in the aerobic reactor recirculation loop $T(SL)_r$
- 1 in the feed sludge line to the aerobic reactor $T(SL)_{in}$
- 1 in the air delivery line to the aerobic reactor $T(AIR)_{in}$
- 1 in the anaerobic digester $T(SL)_d$

The temperatures from the thermometers placed in the recirculation loop were recorded by chart recorder. Guages were used to measure temperatures from the other thermometers with the operating staff manually recording temperatures every half hour on specially designed log sheets.

2a.5 Compilation of Process Operational, Chemical and Bacteriological Data

All the operational, chemical and bacteriological data collected during phase I is contained in the following Tables (Table 2a.1 to Table 2a.9). The data collected from the laboratory scale study is contained in Appendix 3.

Table 2a.1 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge												Aerobic Sludge												Anaerobic Sludge											
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
1	-	19	-	-	-	186	17	43	34	54	191	-	5.8	550	130	-	-	-	22	-	-	-	-	-	-	-	-	-	-	-	13	10	7	9	593	690	7.5	130	2780	-	-	-	-
2	780	20	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	26	40	28	48	183	60	6.3	384	220	-	-	-	14	-	-	-	-	-	-	-	-	-	-	-
3	780	20	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	33	35	27	45	145	98	6.7	218	350	-	-	-	14	12	7	13	561	686	7.6	170	2680	-	-	-	-
4	780	16	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	38	34	26	56	161	85	6.9	170	180	-	-	-	14	11	8	12	562	679	7.4	110	2730	-	-	-	-
5	780	18	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	41	33	26	53	181	98	7.1	220	340	-	-	-	14	14	10	15	590	717	7.4	130	2770	-	-	-	-
6	780	20	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	44	32	25	48	183	85	7.2	125	380	-	-	-	14	14	10	16	595	729	7.3	120	2800	-	-	-	-
7	780	21	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	44	30	22	36	196	163	7.2	144	330	-	-	-	15	14	10	16	566	635	7.5	130	2840	-	-	-	-
8	780	21	-	-	-	50	19	38	30	50	228	-	5.8	720	210	-	-	-	41	29	23	44	209	163	7.2	176	450	-	-	-	15	16	11	14	577	605	7.2	140	2870	-	-	-	-
9	580	25	-	-	-	46	20	41	29	56	229	-	5.7	840	180	-	-	-	40	32	25	39	228	163	7.1	254	490	-	-	-	17	14	10	17	583	645	7.2	160	2800	-	-	-	-
10	480	18	-	-	-	56	18	40	33	58	242	-	5.6	970	120	-	-	-	39	32	25	62	253	163	6.9	570	640	-	-	-	19	15	11	16	568	635	7.2	140	2750	-	-	-	-
11	490	16	-	-	-	48	18	35	29	48	201	-	5.7	660	160	-	-	-	38	33	26	44	257	163	6.5	701	540	-	-	-	21	16	11	17	548	655	7.2	150	2710	-	-	-	-
12	770	21	-	-	-	52	20	46	37	76	180	-	5.8	640	110	-	-	-	39	30	24	45	271	223	7.0	515	730	-	-	-	22	14	10	26	553	605	7.2	170	2390	-	-	-	-
13	780	22	-	-	-	48	21	38	31	61	186	-	5.8	550	210	-	-	-	42	32	26	48	221	185	7.3	260	690	-	-	-	23	19	13	23	546	609	7.2	190	2440	-	-	-	-
14	780	19	-	-	-	44	18	38	31	76	191	-	5.9	570	270	-	-	-	44	31	25	42	220	223	7.4	265	700	-	-	-	24	17	12	26	543	629	7.0	190	2460	-	-	-	-
15	770	21	-	-	-	50	20	39	32	54	213	-	5.7	650	310	-	-	-	46	31	23	34	225	210	7.3	308	570	-	-	-	24	16	11	22	393	619	7.1	190	2550	-	-	-	-
16	630	20	-	-	-	42	20	35	28	54	240	-	5.8	770	520	-	-	-	48	29	23	46	239	210	7.0	280	710	-	-	-	25	28	19	31	543	604	7.1	220	2500	-	-	-	-
17	540	20	-	-	-	54	21	34	27	54	190	-	5.9	350	430	-	-	-	51	28	22	37	213	158	7.2	160	470	-	-	-	26	28	19	29	528	641	7.2	150	2610	-	-	-	-
18	450	17	-	-	-	42	17	44	35	70	206	-	5.7	890	30	-	-	-	51	27	21	34	225	193	7.2	200	400	-	-	-	27	19	14	21	540	622	7.3	180	2560	-	-	-	-
19	480	19	-	-	-	50	18	41	34	56	152	-	5.9	490	60	-	-	-	50	27	21	31	222	212	7.4	140	410	-	-	-	27	27	19	28	540	633	7.2	140	2660	-	-	-	-
20	420	23	-	-	-	46	21	35	29	69	205	-	5.8	520	320	-	-	-	50	28	22	36	194	198	7.2	180	420	-	-	-	28	19	13	22	552	626	7.3	130	2630	-	-	-	-
21	340	21	-	-	-	18	20	40	33	72	208	-	5.9	580	270	-	-	-	51	26	21	30	213	135	7.4	230	380	-	-	-	29	20	14	18	524	624	7.2	100	2690	-	-	-	-
22	330	20	-	-	-	38	19	35	28	46	198	-	5.9	520	250	-	-	-	51	26	21	34	218	185	7.4	290	470	-	-	-	29	16	11	23	539	628	7.3	100	2750	-	-	-	-
23	330	17	-	-	-	8	19	24	19	48	230	-	5.7	780	160	-	-	-	51	24	19	21	229	192	7.5	190	620	-	-	-	29	16	11	23	556	629	7.2	100	2650	-	-	-	-
24	340	17	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	53	21	16	25	255	233	7.9	120	630	-	-	-	29	-	-	-	-	-	-	-	-	-	-	-	-
25	350	21	-	-	-	14	22	38	31	31	209	-	5.7	780	60	-	-	-	51	25	19	24	253	217	7.9	120	590	-	-	-	29	18	12	20	552	640	7.2	100	2840	-	-	-	-
26	330	21	-	-	-	20	20	35	28	51	188	-	6.0	650	110	-	-	-	50	35	28	28	250	208	7.9	110	590	-	-	-	29	-	-	-	-	-	-	-	-	-	-	-	-
27	340	18	-	-	-	28	18	33	26	40	265	127	6.0	730	190	-	-	-	48	25	19	28	271	185	7.7	160	580	-	-	-	28	19	13	21	580	675	7.2	110	2840	-	-	-	-
28	340	21	-	-	-	22	19	37	29	45	266	116	5.9	740	260	-	-	-	47	26	20	26	274	176	7.6	230	490	-	-	-	28	26	18	25	589	620	7.5	110	2750	-	-	-	-
Steady State Period 1 Begins																																									
29	330	20	-	-	-	30	20	44	35	66	277	124	5.7	980	120	-	-	-	49	26	20	37	280	198	7.4	330	480	-	-	-	28	29	20	32	609	673	7.4	160	2840	-	-	-	-
30	320	20	-	-	-	30	20	42	33	59	284	146	5.7	870	280	-	-	-	50	27	21	31	287	209	7.3	350	570	-	-	-	29	37	24	38	617	660	7.2	110	3010	-	-	-	-
31	320	19	-	-	-	28	20	37	28	49	249	116	5.9	730	240	-	-	-	51	27	21	32	290	203	7.4	350	630	-	-	-	29	19	13	24	612	710	7.3	110	2990	-	-	-	-
32	320	21	-	-	-	28	20	39	31	53	221	113	5.9	610	200	-	-	-	51	28	21	36	301	211	7.6	270	680	-	-	-	29	16	11	22	610	765	7.3	110	2970	-	-	-	-
33	330	21	52	15.8	3.8	26	20	43	35	67	230	122	5.8	590	210	-	-	-	52	26	20	34	304	245	7.6	230	660	-	-	-	29	25	17	29	610	642	7.4	100	2950	-	-	-	-
34	320	22	53	15.5	4.1	32	20	40	33	58	249	134	5.6	550	310	-	-	-	53	28	21	32	317	267	7.6	200	740	-	-	-	29	17	12	19	600	648	7.2	100	2900	-	-	-	-
35	320	22	53	15.0	4.3	30	20	40	33	54	218	104	5.9	610	170	-	-	-	54	27	21	35	324	286	7.8	220	690	-	-	-	30	19	13	20	600	645	7.2	130	2730	-	-	-	-
36	290	20	-	-	-	26	21	38	30	50	244	114	5.9	640	230	-	-	-	55	22	17	24	340	303	7.8	220	820	-	-	-	31	33	23	36	605	688	7.2	100	2860	-	-	-	-
37	270	21	55	14.8	4.6	24	22	35	28	50	266	134	5.9	780	230	-	-	-	55	22	17	26	349	304	7.8	260	840	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	-
38	240	24	-	-	-	30	22	37	29	52	237	112	5.8	640	210	-	-	-	55	21	15	24	345	294	7.7	330	750	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	-

Table 2a.2 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge													Aerobic Sludge													Anaerobic Sludge												
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
39	240	23	-	-	-	28	21	38	31	57	228	120	5.8	620	180	-	-	-	56	22	16	27	325	272	7.7	380	640	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	-
40	220	22	56	14.9	4.5	30	21	42	34	60	227	100	5.9	620	160	-	1068	35	56	21	16	25	323	272	7.6	360	640	-	1068	0	31	26	17	28	619	708	7.2	100	2890	-	1511	21	-
41	240	20	56	14.5	4.6	29	19	38	31	80	226	99	5.7	560	180	-	-	-	56	20	15	26	320	259	7.5	350	650	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	-
42	210	18	57	14.4	4.5	25	19	41	33	61	248	107	5.7	800	80	5.0e8	1699	49	56	22	17	31	316	258	7.5	410	600	2.2e5	1000	0	32	-	-	-	-	-	-	-	-	7.2e3	2154	10	-
43	220	20	56	14.2	5.3	26	19	38	31	54	234	82	5.8	700	110	-	1129	63	57	25	19	32	320	272	7.6	390	650	3.0e5	1455	0	32	22	15	25	603	660	7.6	100	2800	1.2e6	2252	15	-
44	210	16	56	14.2	5.1	28	18	41	31	55	244	118	5.6	720	190	-	-	-	56	22	16	27	320	277	7.6	330	720	-	-	-	31	30	21	33	609	672	7.5	110	2840	-	-	-	-
45	210	19	-	-	-	32	19	46	35	57	221	84	5.7	630	180	-	-	-	56	21	16	26	320	275	7.6	350	680	-	-	-	31	30	21	32	608	728	7.6	100	2810	-	-	-	-
46	210	21	-	-	-	33	20	36	29	48	174	88	5.9	340	120	-	-	-	56	23	17	27	317	284	7.7	300	710	-	-	-	31	22	15	23	610	682	7.6	110	2800	-	-	-	-
Steady State Period 1 Ends																			Average Values for the Last 10 days of Period																								
-	227	20	56	14.5	4.8	28	20	39	31	57	230	104	5.8	641	164	-	-	-	56	22	16	27	326	277	7.6	346	688	-	-	-	31	26	18	28	610	690	7.5	104	2828	-	-	-	-
47	220	25	-	-	-	36	21	44	35	57	198	94	5.7	510	140	1.8e8	1669	44	55	23	17	27	315	278	7.6	310	700	3.6e5	1423	0	31	19	13	21	603	692	7.6	100	2840	4.0e3	1661	14	-
48	200	21	-	-	-	32	20	30	25	38	233	119	5.9	570	220	6.0e7	635	52	55	17	12	20	309	257	7.7	320	680	4.0e5	-	-	30	15	10	20	588	703	7.7	120	2880	1.1e4	1015	10	-
49	210	24	-	-	-	42	22	31	25	42	240	143	6.1	620	230	1.8e8	1042	63	55	22	16	21	312	271	7.7	370	670	1.7e5	1266	0	30	24	16	28	600	663	7.4	110	2900	5.2e4	1609	11	-
50	200	21	55	12.5	6.2	45	21	36	30	56	238	109	5.8	620	140	1.6e8	-	-	55	23	18	30	316	215	7.5	310	790	5.8e5	-	-	31	33	21	45	613	933	7.4	100	2800	-	-	-	-
51	200	21	55	12.5	6.3	44	21	37	30	54	264	150	5.6	690	310	-	-	-	56	17	13	19	329	261	7.3	540	260	-	-	-	31	23	17	25	536	631	7.2	210	2400	-	-	-	-
52	220	20	-	-	-	45	22	37	30	47	216	100	5.7	560	230	-	-	-	58	20	15	15	334	298	7.4	400	900	-	-	-	31	30	20	24	577	692	7.4	110	2790	-	-	-	-
53	240	23	-	-	-	42	22	30	25	41	200	100	5.9	420	230	-	-	-	59	23	17	16	341	317	7.6	400	950	-	-	-	31	23	16	24	568	682	7.6	130	2720	-	-	-	-
54	250	21	-	-	-	45	21	34	28	51	226	109	5.7	420	210	1.6e8	1415	47	56	26	20	37	364	361	7.4	370	980	7.0e5	1354	0	31	24	16	24	578	691	7.8	100	2780	-	-	-	-
55	230	23	-	-	-	51	22	42	35	57	251	137	5.6	520	240	-	-	-	52	29	22	32	344	343	7.1	430	770	-	-	-	31	24	17	27	597	683	7.4	100	2840	-	-	-	-
56	240	25	-	-	-	48	23	39	32	61	298	210	5.6	560	430	4.2e7	-	-	48	32	25	39	393	391	6.8	910	480	1.2e6	-	-	31	16	11	22	584	584	7.6	110	2890	2.8e4	-	-	-
57	240	27	-	-	-	49	24	46	39	54	223	86	5.4	440	320	4.2e7	-	-	45	34	27	45	399	443	6.7	1020	490	1.3e6	-	-	31	28	19	31	571	703	7.7	90	2760	-	-	-	-
58	240	23	-	-	-	47	22	44	36	52	237	101	5.3	640	160	-	-	-	42	36	29	46	411	439	6.7	1080	460	-	-	-	31	22	15	33	584	768	7.6	110	2820	-	-	-	-
59	240	25	-	-	-	48	23	38	32	58	217	86	5.2	620	70	5.3e7	-	-	41	36	29	59	408	617	6.6	1070	480	9.2e6	-	-	31	23	16	17	592	728	7.3	110	2810	-	-	-	-
60	250	22	-	-	-	50	22	39	32	58	202	88	5.3	620	30	-	-	-	40	33	27	60	386	594	6.7	1050	350	-	-	-	31	22	15	24	589	672	7.3	120	2780	-	-	-	-
61	450	23	-	-	-	48	20	39	33	52	211	104	5.3	640	50	-	-	-	41	34	28	49	331	354	6.7	560	630	-	-	-	30	32	22	35	534	701	7.3	110	2350	-	-	-	-
62	450	21	-	-	-	48	20	40	34	57	202	80	5.5	540	160	-	-	-	44	33	27	46	308	345	7.3	440	760	-	-	-	30	27	19	30	599	709	7.3	120	2680	-	-	-	-
63	240	22	-	-	-	49	22	39	33	57	215	84	5.5	640	80	-	211	67	42	33	27	36	340	379	7.1	760	550	-	360	0	30	27	19	29	614	736	7.3	120	2880	-	-	-	-
64	600	23	-	-	-	48	21	38	32	54	195	71	5.7	490	110	-	110	25	43	32	27	35	330	356	7.3	530	720	-	-	-	30	29	21	34	553	647	7.3	140	2430	-	2865	5	-
65	600	26	-	-	-	48	23	37	30	55	258	105	5.2	900	50	-	-	-	46	27	22	35	324	335	7.1	390	1010	-	-	-	29	25	18	28	615	726	7.2	160	3060	-	-	-	-
66	610	26	-	-	-	48	23	33	27	44	237	111	5.4	680	120	-	-	-	49	26	21	32	331	511	7.4	290	1120	-	-	-	29	24	17	27	607	697	7.3	150	2850	-	-	-	-
67	610	26	-	-	-	48	23	36	30	53	192	94	5.6	430	170	-	-	-	50	28	23	38	317	351	7.6	220	1030	-	-	-	30	22	15	26	614	739	7.2	160	2930	-	-	-	-
68	640	23	52	17.9	2.3	48	21	34	28	46	197	96	5.6	510	90	6.4e8	578	60	51	27	22	32	308	316	7.7	210	980	1.2e6	410	0	30	26	18	29	603	732	7.3	130	2900	2.1e5	464	11	-
69	640	24	52	17.6	2.6	52	22	40	33	63	203	90	5.7	550	100	2.7e6	338	63	52	29	23	40	310	317	7.8	190	920	2.7e5	491	0	30	28	20	34	564	663	7.3	140	2580	1.8e5	833	10	-
70	630	24	52	17.6	2.7	49	22	40	32	51	285	237	6.0	600	400	2.2e8	367	35	52	31	26	31	324	312	7.8	220	920	2.0e6	462	0	30	25	21	32	538	599	7.1	250	2160	2.2e5	653	13	-
71	760	24	-	-	-	47	22	42	35	60	233	115	5.6	670	150	2.4e9	-	-	53	33	26	39	313	295	7.8	180	910	1.4e6	-	-	30	28	20	38	617	744	7.3	140	2870	1.3e5	-	-	-
72	760	24	54	17.7	2.5	48	23	51	42	61	314	234	5.6	780	320	-	-	-	54	28	22	30	311	272	7.5	250	840	-	-	-	30	27	19	28	559	622	7.3	110	2640	-	-	-	-
73	760	28	-	-	-	48	24	52	44	76	282	206	5.4	790	260	-	-	-	55	31	25	30	324	307	7.5	270	860	-	-	-	31	35	25	36	624	763	7.2	110	2990	-	-	-	-
74	760	30	-	-	-	50	24	54	46	75	238	143	5.4	750	130	-	-	-	56	34	28	43	330	375	7.7	260	940	-	-	-	32	40	28	41	626	743	7.2	130	3030	-	-	-	-

Table 2a.3 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge												Aerobic Sludge												Anaerobic Sludge											
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	C	C	%	%	m^3/d	C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
75	630	25	-	-	-	48	23	54	46	87	-	122	5.4	570	210	8.4e9	475	80	55	39	32	55	-	425	7.7	320	560	8.0e5	259	0	32	25	18	24	-	738	7.3	100	3060	-	-	-	-
76	630	23	51	17.9	1.7	43	21	56	46	-	-	-	5.6	-	-	-	714	44	53	37	30	-	-	-	7.9	-	-	-	359	0	32	27	19	-	-	-	7.4	-	-	-	-	-	-
77	630	23	50	18.0	1.8	49	20	60	49	-	-	-	5.4	-	-	-	-	-	52	43	34	-	-	-	8.0	-	-	-	-	-	32	24	17	-	-	-	7.6	-	-	-	-	-	-
Steady State Period 2 Begins																																										
78	630	23	46	18.2	1.8	49	22	51	42	-	-	-	5.4	-	-	-	-	-	50	41	33	-	-	-	7.9	-	-	-	-	-	31	21	15	-	-	-	7.8	-	-	-	-	-	-
79	630	25	-	-	-	49	22	49	42	-	-	-	5.3	-	-	-	-	-	50	34	28	-	-	-	7.3	-	-	-	-	-	32	34	24	-	-	-	7.3	-	-	-	-	-	-
80	630	22	-	-	-	50	22	49	41	-	-	-	5.5	-	-	-	-	-	50	30	25	-	-	-	7.2	-	-	-	-	-	31	24	17	-	-	-	7.4	-	-	-	-	-	-
81	630	22	-	-	-	49	22	44	37	-	-	-	5.6	-	-	-	-	-	50	34	28	-	-	-	7.3	-	-	-	-	-	31	25	17	-	-	-	7.5	-	-	-	-	-	-
82	630	25	-	-	-	49	23	45	38	-	-	-	5.6	-	-	-	-	-	49	41	34	-	-	-	7.7	-	-	-	-	-	31	22	15	-	-	-	7.4	-	-	-	-	-	-
83	630	23	-	-	-	48	22	46	38	-	-	-	5.7	-	-	-	-	-	49	34	28	-	-	-	7.4	-	-	-	-	-	30	22	15	-	-	-	7.5	-	-	-	-	-	-
84	630	24	-	-	-	50	22	45	37	-	-	-	5.7	-	-	1.0e8	-	-	48	39	32	-	-	-	7.5	-	-	1.0e6	-	-	30	28	19	-	-	-	7.4	-	-	-	-	-	-
85	630	26	-	-	-	51	24	45	36	-	-	-	5.7	-	-	-	-	-	48	37	30	-	-	-	7.5	-	-	-	-	-	30	27	19	-	-	-	7.3	-	-	-	-	-	-
86	630	23	-	-	-	50	21	51	42	-	-	-	5.4	-	-	-	-	-	48	36	29	-	-	-	7.1	-	-	-	-	-	30	25	17	-	-	-	7.3	-	-	-	-	-	-
87	630	23	-	-	-	50	23	52	42	-	-	-	5.5	-	-	-	-	-	49	38	31	-	-	-	7.3	-	-	-	-	-	31	25	17	-	-	-	7.4	-	-	-	-	-	-
88	620	24	-	-	-	51	22	51	42	-	-	-	5.5	-	-	-	-	-	48	39	32	-	-	-	7.5	-	-	-	-	-	31	29	20	-	-	-	7.5	-	-	-	-	-	-
89	630	25	46	18.5	1.6	52	23	47	38	-	-	-	5.2	-	-	-	-	-	48	39	31	-	-	-	6.9	-	-	-	-	-	31	31	22	-	-	-	7.3	-	-	-	-	-	-
90	640	23	-	-	-	50	20	50	41	-	-	-	5.1	-	-	-	-	-	48	38	30	-	-	-	6.5	-	-	-	-	-	30	24	17	-	-	-	7.7	-	-	-	-	-	-
91	640	24	43	18.6	1.8	48	23	56	47	-	-	-	5.6	-	-	-	-	-	46	40	32	-	-	-	7.2	-	-	-	-	-	30	33	23	-	-	-	7.4	-	-	-	-	-	-
92	640	24	44	18.7	1.7	50	23	55	45	-	-	-	5.6	-	-	7.3e8	255	71	47	43	35	-	-	-	7.3	-	-	2.8e7	437	0	30	30	21	-	-	-	7.4	-	-	3.8e4	833	2	-
93	640	26	45	18.6	1.9	49	24	52	45	-	-	-	5.3	-	-	-	-	-	47	43	35	-	-	-	7.3	-	-	-	-	-	30	28	20	-	-	-	7.3	-	-	-	-	-	-
94	660	23	-	-	-	49	20	58	48	-	-	-	5.3	-	-	-	-	-	47	43	35	-	-	-	7.4	-	-	-	-	-	30	24	17	-	-	-	7.5	-	-	-	-	-	-
95	660	23	-	-	-	48	22	55	46	-	-	-	5.4	-	-	-	-	-	47	43	35	-	-	-	7.6	-	-	-	-	-	30	28	20	-	-	-	7.4	-	-	-	-	-	-
96	650	28	45	18.5	1.9	49	24	62	53	100	258	154	5.3	780	90	2.0e8	480	22	47	44	36	61	448	623	7.5	560	1310	3.5e6	312	0	30	23	16	30	654	788	7.5	110	3080	1.0e5	958	1	-
97	660	25	43	18.6	1.8	48	23	58	49	98	238	122	5.3	740	70	-	-	-	47	46	38	65	456	623	7.5	620	1400	-	-	-	30	26	18	31	650	788	7.4	110	3200	-	-	-	-
98	660	25	44	18.5	1.7	51	23	57	48	92	218	111	5.4	710	20	-	288	61	47	47	39	69	457	650	7.6	660	1370	-	506	6	30	29	21	34	640	794	7.3	120	3190	-	122	15	-
99	650	27	-	-	-	48	23	59	50	92	266	153	5.2	1000	0	1.9e7	283	35	47	47	39	72	471	666	7.5	680	1370	3.4e7	553	7	30	22	15	26	666	793	7.3	130	3170	9.0e5	203	19	-
100	660	25	43	18.6	1.6	49	23	55	45	83	245	136	5.4	850	0	-	-	-	47	-	-	-	440	620	7.6	630	1350	-	-	-	30	19	14	20	628	773	7.4	90	3000	-	-	-	-
101	660	24	-	-	-	49	23	50	41	72	230	122	5.3	730	70	-	-	-	47	44	36	60	433	615	7.7	490	1350	-	-	-	30	23	17	18	614	793	7.5	120	3040	-	-	-	-
102	660	25	-	-	-	50	23	47	39	63	243	170	5.5	690	90	-	-	-	47	43	36	63	422	582	7.7	430	1280	-	-	-	29	24	17	31	641	784	7.4	140	3130	-	-	-	-
Steady State Period 2 Ends																			Average Values for the Last 10 days of Period																							
-	656	25	44	18.6	1.8	49	23	55	46	86	242	138	5.3	786	48	-	-	-	47	44	36	65	447	626	7.5	581	1347	-	-	-	30	25	18	27	642	788	7.4	117	3116	-	-	-	-
103	740	27	-	-	-	49	25	50	41	74	223	122	5.3	490	220	9.3e8	260	26	48	44	36	60	434	575	7.5	420	1280	4.3e7	377	10	29	32	24	44	581	715	7.4	200	2420	5.3e5	310	5	-
104	750	27	-	-	-	49	24	53	44	78	221	106	5.3	740	0	5.5e9	215	42	49	42	34	59	404	521	7.4	460	1180	2.0e7	708	4	29	33	24	38	680	798	7.3	120	3280	1.0e4	904	24	-
105	750	26	-	-	-	49	24	45	37	63	245	96	5.4	780	0	-	-	-	50	37	30	31	384	471	7.3	420	1100	-	-	-	30	26	19	30	669	798	7.5	110	3290	-	-	-	-
106	750	25	-	-	-	51	23	46	39	84	221	96	5.3	680	20	5.0e6	-	-	50	43	35	38	365	399	7.3	450	990	4.0e4	-	-	30	24	17	22	675	758	7.2	100	3150	-	-	-	-
107	750	24	-	-	-	52	22	43	35	61	267	117	5.2	1020	0	-	-	-	51	32	26	37	387	453	7.3	380	1360	-	-	-	30	14	10	12	659	735	7.2	150	3290	-			-

Table 2a.4 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge												Aerobic Sludge												Anaerobic Sludge											
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
108	750	25	-	-	-	49	23	34	28	61	238	102	5.4	880	0	-	-	-	-	28	23	26	370	422	7.5	310	1230	-	-	-	30	10	7	7	649	754	7.3	110	3290	-	-	-	-
109	750	25	-	-	-	63	22	28	23	44	250	199	6.0	440	460	-	-	-	48	33	26	31	330	360	7.6	240	1130	-	-	-	31	24	18	24	600	678	7.2	130	2900	-	-	-	-
110	760	26	46	18.5	1.9	49	23	34	28	35	245	135	5.9	570	420	-	-	-	49	29	23	25	352	265	7.4	190	1270	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	-
111	750	25	48	18.4	2.1	50	23	40	33	53	228	87	5.3	810	0	-	-	-	50	33	27	49	288	195	7.5	200	900	-	-	-	31	22	16	19	657	768	7.4	110	3220	-	-	-	-
112	750	25	49	18.2	2.0	50	23	34	28	50	237	112	5.4	860	0	-	-	-	50	27	22	33	277	243	7.4	230	870	-	-	-	31	22	15	22	630	783	7.4	100	3460	-	-	-	-
113	750	27	50	18.0	2.3	49	24	39	32	56	256	114	5.3	820	10	-	-	-	51	32	26	40	333	331	7.5	290	1020	-	-	-	31	18	13	18	651	750	7.4	90	3110	-	-	-	-
114	760	26	50	17.9	2.2	49	24	47	38	60	306	147	5.1	1380	0	-	-	-	52	32	26	37	388	437	7.6	360	1310	-	-	-	31	17	12	16	634	703	7.2	100	3180	-	-	-	-
115	750	25	-	-	-	48	23	26	21	36	288	150	5.4	1010	10	-	-	-	53	23	18	28	387	413	7.7	310	1200	-	-	-	31	23	16	22	660	803	7.3	100	3360	-	-	-	-
116	750	28	-	-	-	48	25	24	20	32	221	97	5.4	630	0	-	-	-	52	30	24	36	354	349	7.6	240	1050	-	-	-	32	21	15	20	654	783	7.4	100	3310	-	-	-	-
Steady State Period 3 Begins																																										
117	760	26	51	18.0	2.3	50	24	35	30	43	236	124	5.4	930	0	-	-	-	51	25	20	28	327	290	7.5	230	940	-	-	-	32	-	-	-	-	-	-	-	-	-	-	-	-
118	750	28	-	-	-	44	26	47	39	56	279	142	5.3	1120	0	-	-	-	51	26	22	27	308	249	7.4	260	680	-	-	-	32	18	13	15	653	778	7.4	120	3120	-	-	-	-
119	750	25	49	18.1	2.2	52	22	28	24	30	258	131	5.3	1000	0	-	-	-	50	26	21	28	296	227	7.0	360	570	-	-	-	32	25	18	28	617	803	7.7	140	3050	-	-	-	-
120	750	29	-	-	-	49	26	38	31	47	254	133	5.3	1080	0	-	-	-	50	37	30	42	307	244	6.9	360	610	-	-	-	32	38	27	48	641	780	7.4	140	3090	-	-	-	-
121	750	26	47	18.5	1.8	48	24	43	36	57	336	162	5.1	1480	0	-	-	-	50	33	27	34	309	333	6.7	730	620	-	-	-	32	27	20	37	624	786	7.5	150	3140	-	-	-	-
122	760	25	-	-	-	48	24	48	41	71	414	156	5.2	1020	280	-	-	-	50	32	26	31	289	479	7.5	500	1300	-	-	-	32	-	-	-	-	-	-	-	-	-	-	-	-
123	760	25	-	-	-	48	23	43	36	66	421	119	5.3	680	30	-	-	-	50	33	27	35	228	511	7.7	500	1300	-	-	-	32	-	-	-	-	-	-	-	-	-	-	-	-
124	750	27	46	18.4	2.0	49	26	44	36	59	235	125	5.5	960	0	-	-	-	50	29	23	29	412	511	8.0	590	1410	-	-	-	32	-	-	-	-	-	-	-	-	-	-	-	-
125	760	30	-	-	-	46	26	48	39	75	226	100	5.5	910	0	-	-	-	50	35	29	38	416	525	8.0	580	1230	-	-	-	33	-	-	-	-	-	-	-	-	-	-	-	-
126	750	27	-	-	-	59	26	40	33	55	233	103	5.4	1070	0	-	-	-	50	30	25	31	396	478	7.9	480	1260	-	-	-	33	-	-	-	-	-	-	-	-	-	-	-	-
127	750	27	-	-	-	49	26	41	34	53	211	80	5.3	980	0	-	-	-	50	37	30	35	378	447	7.9	350	1210	-	-	-	33	-	-	-	-	-	-	-	-	-	-	-	-
128	750	23	49	18.0	2.1	49	23	41	33	46	234	81	5.2	940	0	-	-	-	50	39	31	32	367	423	7.3	430	1210	-	-	-	34	-	-	-	-	-	-	-	-	-	-	-	-
129	750	27	-	-	-	48	24	36	29	48	226	79	5.2	930	0	-	-	-	51	27	22	33	367	369	7.6	340	1120	-	-	-	34	-	-	-	-	-	-	-	-	-	-	-	-
130	760	26	-	-	-	48	24	32	26	66	189	67	5.4	730	0	-	-	-	51	26	21	32	338	589	7.3	340	910	-	-	-	34	-	-	-	-	-	-	-	-	-	-	-	-
131	760	25	49	18.4	2.0	48	24	41	34	67	194	74	5.4	830	0	1.0e10	297	15	51	25	20	22	304	233	7.7	290	750	1.2e7	341	0	33	-	-	-	-	-	-	-	-	-	-	-	-
132	750	26	-	-	-	49	25	42	34	60	198	53	5.3	1340	0	8.0e9	348	33	51	-	-	-	-	-	-	-	-	4.0e6	404	0	33	24	17	33	680	833	7.5	220	3170	4.7e4	137	12	-
133	750	25	48	18.4	2.0	48	24	36	29	38	154	79	5.4	760	0	-	213	61	51	26	21	28	212	195	6.9	210	500	-	141	0	34	20	14	24	531	833	7.5	90	3310	-	580	0	-
134	750	24	46	18.5	1.9	47	23	32	26	34	149	49	5.3	690	0	-	-	-	50	27	22	43	208	175	7.5	150	470	-	-	-	34	24	17	28	549	815	7.7	90	3220	-	-	-	-
135	750	25	-	-	-	48	25	32	26	49	151	59	4.8	730	0	-	-	-	50	20	16	26	206	155	7.1	200	400	-	-	-	34	22	15	31	518	783	7.5	130	3140	-	-	-	-
136	750	25	-	-	-	48	26	29	23	33	161	55	5.3	610	0	-	-	-	50	26	20	32	201	145	7.2	120	490	-	-	-	34	21	15	24	525	790	7.5	100	3200	-	-	-	-
137	750	24	-	-	-	48	23	30	24	40	151	63	5.0	600	0	-	-	-	49	21	17	32	199	133	7.4	60	470	-	-	-	34	21	15	29	546	776	7.4	100	3180	-	-	-	-
138	750	22	46	18.4	2.0	47	22	37	29	44	198	83	5.0	1280	0	5.2e9	141	33	49	23	18	24	280	185	6.9	410	500	1.0e6	221	0	34	25	18	27	550	763	7.5	120	3180	1.0e3	207	4	-
139	760	22	-	-	-	48	20	33	26	46	188	79	5.2	930	0	-	-	-	49	25	19	32	253	195	7.3	170	690	-	-	-	34	20	14	21	460	708	7.5	120	3150	-	-	-	-
140	780	22	48	18.3	2.2	48	22	35	27	47	178	64	5.2	950	0	4.8e9	-	-	50	22	17	24	274	205	7.3	150	560	1.0e5	-	-	33	24	17	30	530	777	7.5	100	3150	-	-	-	-
141	770	22	47	18.4	2.0	48	22	36	29	53	169	76	5.2	910	0	-	168	31	50	29	22	32	246	210	7.6	140	530	-	78	0	34	20	14	24	602	748	7.5	120	3050	-	300	19	-
Steady State Period 3 Ends																			Average Values for the Last 10 days of Period																							
-	756	24	47	18.4	2.0	48	23	34	27	44	170	66	5.2	880	0	-	-	-	50	24	19	30	231	178	7.2	179	512	-	-	-	34	22	16	27	549	783	7.5	119	3175	-	-	-	-

Table 2a.5 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge												Aerobic Sludge												Anaerobic Sludge											
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
142	770	23	-	-	-	58	21	32	26	45	226	118	5.2	1220	0	-	-	-	51	26	20	38	284	224	7.7	260	590	-	-	-	34	18	12	13	632	688	7.6	180	3040	-	-	-	-
143	770	23	-	-	-	49	22	32	26	45	210	95	5.2	960	0	-	-	-	52	28	22	30	299	234	7.3	260	610	-	-	-	34	18	13	17	616	741	7.6	140	3130	-	-	-	-
144	770	22	-	-	-	47	22	31	26	48	194	93	5.2	850	0	-	-	-	53	25	20	34	312	265	7.4	200	740	-	-	-	34	28	20	29	627	742	7.5	150	3060	-	-	-	-
145	610	24	53	17.5	2.6	49	25	33	27	41	210	123	5.5	850	0	5.0e7	-	-	53	25	19	26	317	278	7.0	460	540	1.8e6	-	-	34	20	14	21	592	711	7.6	130	2820	-	-	-	-
Steady State Period 4 Begins																																										
146	520	25	-	-	-	48	23	31	25	41	229	120	5.4	1060	0	-	-	-	54	24	19	24	310	273	7.9	160	770	6.0e5	147	0	34	24	17	28	617	748	7.7	110	2960	5.0e2	614	10	-
147	510	25	54	16.8	3.3	47	27	31	25	35	285	114	5.2	960	0	-	-	-	54	24	18	18	386	253	7.8	240	780	3.0e5	320	0	34	16	11	13	738	735	7.8	130	2930	1.6e3	593	0	-
148	480	25	54	16.9	3.1	48	25	31	25	38	280	93	5.3	890	0	-	541	52	54	25	20	22	398	259	7.7	200	750	-	169	0	34	13	9	11	763	707	7.6	250	2850	-	685	12	-
149	520	23	-	-	-	53	24	34	28	46	229	111	5.2	970	0	-	-	-	54	23	18	26	288	249	7.6	190	660	-	-	-	34	12	8	11	551	693	7.7	130	2910	-	-	-	-
150	470	23	-	-	-	48	23	33	27	69	200	74	5.1	830	0	-	-	-	54	20	16	28	290	448	7.6	190	620	-	-	-	34	14	10	22	555	723	7.7	110	2850	-	-	-	-
151	490	23	-	-	-	48	23	28	22	38	237	195	6.1	660	180	-	-	-	53	22	18	37	272	221	7.7	190	560	-	-	-	34	13	9	19	557	701	7.8	110	2890	-	-	-	-
152	530	25	53	16.4	3.2	49	23	40	33	68	311	169	5.3	1610	0	-	425	74	52	15	12	25	312	209	7.4	200	600	-	-	-	34	10	7	12	614	684	7.9	110	2990	-	610	4	-
153	460	23	53	16.3	3.5	48	24	36	29	57	255	99	5.3	1020	0	-	-	-	53	19	15	39	298	175	7.0	300	500	-	-	-	34	11	8	12	657	718	7.6	120	2850	-	-	-	-
154	450	25	55	16.1	3.7	50	24	34	27	47	285	158	5.5	970	0	-	111	33	54	23	19	27	278	131	7.2	210	480	-	-	-	34	12	9	7	668	693	7.6	110	2800	-	-	-	-
155	430	26	54	16.0	3.8	48	23	38	31	53	271	101	5.3	1040	0	-	208	40	55	27	22	32	255	129	7.4	180	480	-	-	-	35	13	9	9	667	705	7.5	110	2860	-	566	1	-
156	400	24	55	15.8	3.9	52	23	38	32	53	250	119	5.0	1470	0	-	-	-	55	25	20	26	232	127	6.8	290	360	-	-	-	35	22	15	21	549	678	7.7	110	2700	-	-	-	-
157	400	24	-	-	-	50	23	36	29	59	233	122	5.2	1050	0	-	-	-	56	27	22	36	226	129	7.2	230	330	-	-	-	35	19	14	22	569	728	7.8	110	2790	-	-	-	-
158	390	22	-	-	-	48	23	29	24	46	215	114	5.3	950	0	-	-	-	55	28	22	26	230	137	7.4	180	370	-	-	-	35	18	12	20	577	668	7.8	130	2670	-	-	-	-
159	420	24	55	15.8	3.9	50	24	42	35	54	239	134	5.5	990	0	-	220	26	55	-	-	-	219	122	7.1	280	280	-	-	-	35	13	9	14	563	683	7.6	140	2630	-	333	13	-
Steady State Period 4 Ends																			Average Values for the Last 10 days of Period																							
-	444	24	54	16.1	3.7	49	23	35	29	54	250	128	5.4	1059	18	-	-	-	54	23	18	31	261	183	7.3	225	458	-	-	-	34	14	10	16	598	698	7.7	116	2803	-	-	-	-
160	390	24	57	15.6	4.0	43	24	39	32	60	274	163	5.5	1190	0	-	410	36	57	-	-	-	225	128	7.0	350	240	-	-	-	35	10	7	17	573	688	7.7	120	2620	-	47	3	-
161	380	22	59	15.4	4.2	37	23	35	29	45	250	123	5.4	1080	0	-	151	74	59	-	-	-	296	233	7.2	430	360	-	467	0	35	13	9	7	583	662	7.7	130	2630	-	365	6	-
162	470	25	58	15.6	4.0	43	24	37	32	45	212	97	5.6	820	0	-	91	67	58	29	24	32	276	247	7.6	610	330	-	142	0	35	15	11	15	577	670	7.6	140	2640	-	-	-	-
163	580	26	56	16.6	3.0	48	24	40	34	50	241	135	5.5	1210	0	-	-	-	57	30	24	26	259	199	7.2	560	420	-	-	-	35	15	10	8	566	662	7.3	110	2730	-	-	-	-
164	720	26	-	-	-	48	25	34	28	40	238	125	5.6	1000	0	-	-	-	56	31	25	22	250	201	7.3	330	630	-	-	-	35	19	14	9	590	675	7.3	180	2720	-	-	-	-
165	770	25	-	-	-	49	23	37	31	47	190	94	5.7	660	0	-	-	-	54	27	22	26	268	223	7.6	310	720	-	-	-	35	19	14	14	583	689	7.3	140	2700	-	-	-	-
166	760	26	53	17.9	2.4	46	22	47	39	60	269	170	5.7	1020	0	-	-	-	53	17	14	34	268	268	7.5	300	742	3.0e4	446	0	35	8	6	10	600	678	7.5	130	2610	2.2e4	758	21	-
167	770	23	-	-	-	50	23	48	40	73	258	129	5.6	1450	0	8.4e10	234	20	53	22	18	37	348	333	7.6	320	866	-	85	0	35	7	5	10	604	669	7.5	160	2650	3.0e4	422	12	-
168	770	24	51	17.8	2.4	48	23	43	36	56	239	114	5.6	1070	0	3.4e10	301	30	53	36	29	31	373	380	7.7	400	866	-	341	0	35	13	9	10	605	679	7.4	150	2680	2.6e4	221	21	-
169	770	25	50	17.8	2.4	51	24	42	35	69	208	96	5.7	750	0	2.0e8	357	35	52	23	18	30	395	334	7.9	430	1010	2.2e5	113	0	35	22	16	9	588	673	7.4	130	2660	-	372	18	-
170	770	23	47	18.1	2.2	47	22	49	36	76	284	122	5.2	1400	0	-	-	-	51	34	27	38	440	481	7.5	600	1010	-	-	-	35	15	11	16	626	673	7.2	190	2760	-	-	-	-
171	770	25	-	-	-	48	22	38	31	85	257	107	5.2	1170	0	-	-	-	51	30	24	23	453	697	7.4	590	1060	-	-	-	35	21	15	23	642	734	7.2	190	2810	-	-	-	-
172	770	25	-	-	-	48	22	28	23	60	223	105	5.3	950	0	-	-	-	51	40	32	54	458	545	7.6	590	1070	-	-	-	36	18	13	22	633	690	7.3	150	2710	-	-	-	-

Table 2a.6 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge												Aerobic Sludge												Anaerobic Sludge											
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
Steady State Period 5 Begins																																									
173	770	24	-	-	-	59	22	27	22	32	212	102	6.0	490	0	1.3e11	544	50	49	42	34	30	498	653	7.9	590	1260	3.1e6	359	0	36	21	15	12	648	744	7.4	90	2880	-	-	-	-
174	770	22	-	-	-	59	22	44	36	49	207	104	5.9	410	230	6.2e9	238	24	47	41	34	46	470	625	7.9	510	1240	8.0e4	-	-	36	18	13	11	650	721	7.4	150	2810	8.0e3	542	10	-
175	770	23	42	18.5	1.9	60	20	38	31	51	251	136	5.6	730	10	-	84	31	46	33	27	38	447	559	7.7	550	1150	-	425	12	35	20	15	16	643	770	7.2	100	2840	-	-	-	-
176	770	23	44	18.4	2.0	60	20	48	40	62	252	135	5.5	830	20	-	-	-	46	42	34	48	470	575	7.4	520	1280	-	-	-	35	19	14	14	647	743	7.2	130	2870	-	-	-	-
177	760	25	43	18.5	1.9	60	22	48	40	62	337	151	5.2	1610	0	-	-	-	46	-	-	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-	-	-	-
178	770	21	-	-	-	60	19	50	42	63	296	144	5.3	1330	0	-	-	-	43	42	34	55	504	637	7.7	810	1260	-	-	-	35	22	16	22	643	761	7.6	270	2630	-	-	-	-
179	770	22	-	-	-	60	22	42	34	50	210	111	5.9	630	10	-	-	-	43	49	40	52	498	637	7.6	800	1200	-	-	-	34	18	13	21	640	757	7.6	230	2630	-	-	-	-
180	770	23	42	18.6	1.7	61	21	48	40	82	267	158	5.5	790	0	-	-	-	45	36	29	49	446	531	7.5	580	1140	-	-	-	34	16	11	15	677	763	7.4	110	2990	-	-	-	-
181	760	23	41	18.5	1.7	60	21	49	42	72	245	117	5.6	630	70	-	-	-	44	41	33	52	450	512	7.0	920	680	-	-	-	34	15	11	12	675	736	7.5	140	2960	-	-	-	-
182	770	23	-	-	-	55	20	67	58	82	207	110	5.5	810	0	-	550	32	44	39	32	55	395	461	7.1	920	830	-	758	0	34	16	11	18	652	760	7.4	160	3050	-	556	20	373
183	750	21	40	18.6	1.9	60	19	46	39	65	300	161	5.4	1290	0	-	-	-	43	40	30	52	472	565	7.2	820	1030	-	-	-	33	20	14	25	671	767	7.3	150	3090	-	-	-	-
184	750	23	41	18.5	1.8	62	19	45	38	63	281	144	5.5	1030	0	-	-	-	43	26	21	34	487	593	7.4	890	1050	-	-	-	34	16	11	19	666	763	7.5	120	3080	-	-	-	-
185	750	23	-	-	-	62	21	39	33	53	256	130	5.6	780	0	-	-	-	43	28	22	36	461	541	7.3	920	880	-	-	-	34	17	12	22	662	758	7.6	120	2950	-	-	-	-
186	740	23	-	-	-	60	21	36	31	46	201	100	5.8	590	30	-	-	-	43	33	27	43	460	545	7.4	730	1160	-	-	-	33	18	12	19	649	818	7.5	110	2900	-	-	-	-
Steady State Period 5 Ends																			Average Values for the Last 10 days of Period																							
-	759	23	41	18.5	1.8	60	20	47	40	64	260	133	5.5	949	11	-	-	-	44	37	30	48	464	558	7.4	821	1026	-	-	-	34	18	12	19	659	765	7.5	157	2920	-	-	-	-
187	1200	24	-	-	-	61	22	45	39	65	225	119	5.6	930	0	-	-	-	45	24	21	30	462	557	7.3	800	1050	-	-	-	33	19	14	30	680	780	7.5	280	2880	-	-	-	-
188	1200	24	-	-	-	60	22	45	39	59	235	116	5.7	850	0	-	-	-	48	33	28	37	408	437	7.3	690	960	-	-	-	33	24	17	28	688	761	7.4	340	2800	-	-	-	-
189	1200	21	-	-	-	60	19	45	38	76	204	105	5.6	670	0	-	439	48	50	34	28	50	352	369	7.6	410	1020	-	583	3	33	22	16	23	657	803	7.5	140	3090	1.5e5	554	31	247
190	1200	23	-	-	-	60	20	45	37	92	243	188	5.3	1400	0	-	-	-	50	21	17	36	318	317	7.3	440	840	-	-	-	33	24	17	22	602	770	7.2	250	2730	-	-	-	-
191	1200	22	-	-	-	62	21	28	23	57	245	169	5.4	1120	0	-	-	-	50	33	27	45	302	279	7.3	390	740	-	-	-	33	24	17	26	610	773	7.3	120	3180	-	-	-	-
192	1200	23	-	-	-	61	20	36	29	61	208	170	5.7	860	0	-	-	-	50	26	21	28	311	261	7.4	290	960	-	-	-	34	20	14	22	633	773	7.6	180	2970	-	-	-	-
193	1200	22	-	-	-	60	20	28	24	66	167	133	5.8	510	0	-	-	-	48	27	22	30	316	301	7.5	260	790	-	-	-	34	22	16	32	655	815	7.5	130	3040	-	-	-	-
194	1200	22	-	-	-	60	19	31	26	65	171	132	5.7	600	0	-	-	-	46	37	29	40	323	329	7.5	430	770	-	-	-	34	23	17	22	643	745	7.4	200	3000	-	-	-	-
195	1200	26	-	-	-	60	23	35	29	62	195	145	5.8	650	0	-	-	-	48	35	28	46	325	322	7.1	230	870	-	-	-	34	22	15	21	656	778	7.3	160	3020	-	-	-	-
196	1200	22	44	19.0	1.5	62	19	45	36	74	210	148	5.6	1170	0	6.0e9	265	63	47	31	25	41	325	345	7.4	240	880	1.9e6	964	5	34	20	14	21	640	758	7.3	170	2890	1.0e5	664	66	-
Steady State Period 6 Begins																																									
197	1200	24	44	19.0	1.5	60	21	47	38	76	266	141	5.3	840	10	-	-	-	47	37	30	36	354	333	7.2	340	830	-	-	-	34	20	14	19	674	731	7.2	140	2930	-	-	-	-
198	1200	25	45	19.0	1.3	60	20	47	38	75	307	139	5.4	1430	0	-	-	-	49	35	28	53	391	361	7.2	370	870	-	-	-	33	15	10	21	701	844	7.2	120	3200	-	-	-	-
199	1200	22	-	-	-	60	19	33	27	61	245	100	5.5	960	0	-	-	-	48	37	30	48	403	357	7.0	460	950	-	-	-	33	11	8	15	694	783	7.2	120	3050	-	-	-	-
200	1200	21	-	-	-	60	19	39	30	44	177	58	5.6	660	0	-	-	-	46	33	26	36	364	313	7.3	320	750	-	-	-	32	32	23	16	716	817	7.1	140	3110	-	-	-	-
201	1200	23	42	19.1	1.3	60	18	45	36	56	215	59	5.2	1020	0	-	-	-	44	29	23	25	320	248	7.2	300	570	-	-	-	32	36	25	46	693	838	7.3	370	2810	-	-	-	-
202	1200	23	-	-	-	60	18	55	44	81	224	59	5.3	1010	0	-	-	-	44	37	29	27	268	160	6.5	440	300	-	-	-	32	32	23	35	687	824	7.2	300	2760	-	-	-	-
203	1200	21	40	19.1	1.4	60	18	58	49	71	214	82	5.4	1020	0	3.6e9	206	45	43	36	28	30	226	167	6.4	480	340	7.0e5	308	12	31	31	22	26	633	753	7.3	140	2960	-	889	0	291
204	1200	21	41	19.0	1.5	60	18	59	47	86	211	90	5.5	970	0	-	-	-	44	42	33	45	248	183	6.7	420	410	-	-	-	30	40	28	38	623	808	7.2	160	2990	-	-	-	-

Table 2a.7 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge													Aerobic Sludge													Anaerobic Sludge												
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
205	1200	21	42	19.0	1.6	60	17	48	37	59	302	82	4.9	1970	0	-	-	-	44	41	32	53	287	223	6.5	770	400	-	-	-	31	52	36	54	687	840	7.1	470	2730	-	-	-	
206	1200	22	-	-	-	60	19	44	34	50	254	65	5.0	1510	0	-	-	-	43	44	34	31	304	239	6.5	910	290	-	-	-	32	38	27	51	705	845	7.0	240	3100	-	-	-	
207	1200	20	-	-	-	57	19	40	32	46	195	51	5.1	970	0	-	-	-	44	32	24	42	263	153	6.8	950	200	-	-	-	33	37	26	66	711	818	7.1	170	3080	-	-	-	
208	1200	19	-	-	-	60	18	46	36	50	282	62	5.0	1410	0	-	-	-	43	39	31	48	286	191	6.7	510	480	-	-	-	33	45	30	45	776	823	7.3	210	2940	-	-	-	
209	1200	19	-	-	-	60	18	39	30	48	259	79	5.3	1230	0	-	-	-	44	39	31	52	273	181	6.5	800	100	-	-	-	33	30	21	36	765	783	7.3	130	2880	-	-	-	
210	1200	20	41	19.1	1.3	60	17	41	31	51	268	83	5.2	1290	0	-	-	-	45	40	32	51	255	153	6.8	410	430	-	-	-	33	34	24	38	753	788	7.3	170	2800	-	-	-	368
211	1200	19	41	19.0	1.4	60	16	49	38	56	251	93	5.7	870	0	-	-	-	44	39	31	47	267	182	7.2	290	460	-	-	-	33	14	10	16	700	686	7.5	110	2650	-	-	-	

Steady State Period 6 Ends

Average Values for the Last 10 days of Period

-	1200	20	41	19.0	1.4	60	18	48	38	60	246	75	5.2	1225	0	-	-	-	44	39	30	43	268	183	6.7	598	341	-	-	-	32	35	25	40	704	797	7.2	210	2889	-	-	-	

212	1200	20	40	19.2	1.3	61	18	52	39	70	371	116	5.0	2360	0	-	-	-	43	41	32	42	281	171	7.2	260	390	-	-	-	33	15	10	16	771	793	7.3	650	2460	-	-	-	
213	1200	21	-	-	-	61	19	56	44	75	339	84	5.0	1880	0	-	-	-	42	45	35	54	306	192	6.3	570	330	-	-	-	32	9	6	10	734	748	7.2	190	2710	-	-	-	
214	1200	20	-	-	-	56	18	48	38	73	253	87	5.3	960	0	-	-	-	41	46	35	52	278	165	6.4	540	350	-	-	-	33	8	6	9	733	723	7.1	150	2730	-	-	-	
215	1200	21	-	-	-	56	19	44	34	55	287	123	5.2	1610	0	-	-	-	41	40	31	51	226	162	6.1	850	40	-	-	-	33	7	5	-	639	668	7.3	180	2750	-	-	-	
216	1200	22	-	-	-	60	19	39	31	50	307	130	5.4	1700	0	-	-	-	42	41	32	49	188	89	6.6	520	170	-	-	-	33	46	31	50	663	826	7.2	540	2520	-	-	-	
217	1200	20	40	19.2	1.3	60	19	52	42	72	361	150	5.2	2100	0	-	-	-	42	42	32	69	172	41	6.7	480	110	-	-	-	33	39	27	45	700	838	7.2	320	2820	-	-	-	
218	620	19	-	-	-	30	19	52	41	72	288	108	5.3	1370	0	-	270	42	42	42	33	52	186	44	6.4	540	180	-	469	0	34	36	24	48	691	848	7.1	210	3000	-	-	-	
219	620	21	-	-	-	29	19	53	44	79	338	111	5.0	2190	0	-	-	-	44	40	31	54	244	127	6.2	1020	0	-	-	-	34	36	24	45	703	873	7.1	420	2940	-	-	-	
220	620	20	-	-	-	38	18	44	36	70	290	101	5.1	1710	0	-	-	-	45	39	31	58	253	149	6.2	980	10	-	-	-	34	31	21	22	727	898	7.3	580	2930	-	-	-	
221	620	19	-	-	-	28	18	47	40	63	252	83	5.2	1240	0	-	-	-	48	38	30	46	250	161	6.5	920	170	-	-	-	34	42	29	66	723	868	7.1	420	2930	-	-	-	
222	620	24	49	18.5	1.9	23	20	58	49	90	340	175	5.1	1890	0	-	-	-	50	39	32	40	269	201	6.9	590	420	-	-	-	35	52	35	35	714	658	7.2	210	3240	-	-	-	
223	600	24	-	-	-	26	21	44	36	90	293	128	5.6	1050	50	-	-	-	52	40	32	49	308	285	7.3	430	670	-	-	-	35	52	36	37	735	568	7.1	260	3290	-	-	-	
224	750	24	55	17.9	2.4	24	19	35	28	66	297	110	5.2	1710	0	-	-	-	54	39	31	54	363	381	7.5	560	940	3.0e5	-	-	35	59	38	48	737	803	7.2	210	3270	-	-	-	
225	750	20	53	17.9	2.3	24	18	38	31	78	260	100	5.3	1240	0	-	-	-	54	38	30	55	385	399	7.9	490	1160	-	-	-	35	55	37	40	725	878	7.1	180	3140	-	-	-	
226	750	19	53	18.0	2.3	32	17	39	32	72	274	85	5.1	1330	0	-	-	-	53	27	21	35	420	471	8.1	530	1400	-	-	-	35	55	36	52	724	863	7.1	530	3220	-	-	-	
227	740	19	-	-	-	38	19	44	36	87	247	59	5.1	1100	0	-	-	-	52	32	25	35	426	459	8.1	440	1460	-	-	-	35	52	35	50	742	875	7.2	230	3280	-	-	-	
228	750	17	-	-	-	33	17	52	41	61	244	78	5.4	1010	0	-	-	-	50	33	26	38	431	459	7.8	600	1210	-	-	-	34	52	35	50	724	868	7.2	240	3060	-	-	-	
229	760	18	47	18.5	1.9	36	16	43	34	71	227	69	5.3	1060	0	-	-	-	50	32	26	30	393	409	8.0	470	1060	-	-	-	34	49	33	43	743	888	7.2	220	3240	-	-	-	
230	750	18	-	-	-	36	16	52	42	79	232	77	5.6	910	0	-	-	-	50	34	26	47	379	373	7.7	450	930	-	-	-	33	55	37	57	748	893	7.1	150	3300	-	-	-	
231	750	18	46	18.5	1.9	36	17	55	44	72	250	56	5.1	1190	0	-	-	-	49	38	30	59	350	363	7.7	390	860	-	-	-	33	58	39	57	740	908	7.2	200	3170	-	-	-	
232	750	17	-	-	-	36	16	53	43	93	220	51	5.2	1020	0	-	-	-	49	43	34	56	367	374	7.6	440	840	-	-	-	33	70	46	72	740	968	7.1	260	3140	-	-	-	
233	750	18	-	-	-	25	18	50	42	69	287	52	4.9	1820	0	-	-	-	50	38	31	53	397	409	7.5	500	1080	-	-	-	33	62	43	72	714	988	7.7	260	3210	-	-	-	
234	740	18	-	-	-	24	17	37	31	66	255	51	5.0	1330	0	-	-	-	51	39	32	56	399	443	7.7	410	1140	-	-	-	33	24	18	27	678	832	7.7	200	3070	-	-	-	
235	710	18	-	-	-	24	17	35	29	70	230	182	5.8	840	0	-	-	-	51	37	30	55	431	465	7.8	610	990	-	-	-	32	25	17	27	726	863	7.4	260	2990	-	-	-	
236	690	19	-	-	-	24	18	52	44	77	303	178	5.4	1380	0	-	705	68	51	39	31	48	436	461	7.7	460	1090	-	892	0	32	25	17	25	724	833	7.5	180	3230	-	-	-	
237	660	18	-	-	-	24	17	55	47	71	266	117	5.6	1010	0	3.0e9	-	-	50	39	31	42	422	461	7.7	490	1060	6.0e5	-	-	32	24	17	30	733	873	7.5	130	3270	-	-	-	
238	640	18	-	-	-	24	17	61	51	76	326	154	5.4	1550	0	-	-	-	49	40	32	54	416	449	7.9	500	950	-	-	-	31	25	17	-	733	868	7.5	180	3120	-	-	-	
239	640	18	-	-	-	24	17	58	48	68	296	151	5.6	1330	0	-	-	-	49	55	47	50	429	473	7.8	540	1010	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	
240	620	18	-	-	-	24	16	51	42	90	366	164	5.1	1300	0	-	-	-	50	39	31	55	451	503	8.1	620	1010	-	-	-	31	24	17	29	749	893	7.4	140	3260	-	-	-	
241	630	18	-	-	-	24	16	50	42	80	294	92	5.0	1050	0	-	-	-	49	38	31	61	519	567	7.5	780	1120	-	-	-	31	23	16	27	748	873	7.3	140	3230	-	-	-	

Table 2a.8 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge													Aerobic Sludge													Anaerobic Sludge												
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
242	650	16	-	-	-	24	15	60	50	69	228	74	5.2	800	0	-	-	-	47	44	36	67	527	605	7.6	820	1230	-	-	-	30	25	17	28	749	903	7.2	180	3270	-	-	-	-
243	650	14	-	-	-	24	14	66	53	84	285	150	5.3	1910	0	-	-	-	46	19	16	43	492	653	7.9	990	1010	-	-	-	30	10	7	27	703	929	7.8	330	2980	-	-	-	-
244	760	16	-	-	-	24	14	68	52	78	251	136	5.4	1490	0	-	-	-	47	41	33	50	501	658	8.0	760	1390	-	-	-	29	17	12	28	731	987	7.5	270	3170	-	-	-	-
245	770	16	-	-	-	24	16	66	53	82	244	86	5.3	1480	0	-	-	-	48	33	27	54	490	641	7.9	680	1340	-	-	-	29	21	15	22	731	900	7.6	230	3160	-	-	-	-
Steady State Period 7 Begins																																										
246	770	16	47	18.5	2.0	35	16	60	49	89	219	67	5.4	1260	0	-	-	-	49	38	31	56	452	591	8.1	650	1250	-	-	-	28	24	17	25	739	903	7.4	210	3210	-	-	-	-
247	770	17	47	18.5	1.9	36	16	55	45	82	309	92	4.9	1710	0	-	-	-	49	41	33	51	466	554	7.2	870	1180	-	-	-	28	20	14	28	738	913	7.3	180	3380	-	-	-	-
248	770	17	-	-	-	36	17	52	42	78	298	89	5.0	1550	0	-	-	-	50	40	32	60	441	519	7.2	790	1070	-	-	-	28	18	12	13	747	900	7.2	190	3250	-	-	-	-
249	770	17	-	-	-	36	17	56	46	84	235	72	5.2	1111	0	-	-	-	50	40	33	53	416	502	7.6	660	1090	-	-	-	28	25	17	18	742	898	7.3	170	3230	-	-	-	-
250	770	17	47	18.5	1.9	36	17	63	49	82	232	132	5.6	1180	0	-	-	-	50	38	32	56	412	519	7.8	850	880	-	-	-	28	8	5	28	742	921	7.4	350	3000	-	-	-	-
251	780	17	-	-	-	36	16	66	52	77	260	91	5.4	1370	0	-	-	-	50	37	30	56	434	541	7.7	790	970	-	-	-	28	14	9	27	739	923	7.5	260	3060	-	-	-	-
252	790	18	45	18.5	1.7	37	16	57	46	100	274	64	5.0	1810	0	-	-	-	49	41	33	79	464	572	7.8	870	980	-	-	-	28	6	3	9	760	901	7.3	210	3190	-	-	-	156
253	780	16	45	18.5	1.8	36	16	63	50	79	225	59	5.2	1260	0	1.0e10	80	60	48	47	37	51	461	569	7.8	700	1050	1.0e5	139	0	28	-	-	-	-	-	-	-	-	-	-	-	-	-
254	780	15	44	18.6	1.6	38	15	46	37	60	261	97	5.1	1710	0	-	-	-	48	45	36	52	448	551	7.5	710	1090	-	-	-	28	15	10	13	763	898	7.2	210	3390	-	-	-	-
255	770	15	-	-	-	34	15	51	41	77	224	82	5.2	1330	0	-	-	-	48	44	36	54	406	479	7.6	550	1080	-	-	-	27	30	21	30	766	943	7.2	230	3330	-	-	-	-
256	770	15	-	-	-	36	15	57	47	71	202	62	5.2	1270	0	-	-	-	48	45	36	58	404	445	7.4	670	970	-	-	-	28	15	10	10	763	877	7.2	220	3180	-	-	-	-
257	770	15	46	18.6	1.7	36	16	41	35	62	218	73	5.2	1120	0	-	-	-	48	40	33	54	379	402	7.6	620	930	-	-	-	29	9	6	11	766	903	7.4	170	3330	-	-	-	-
258	770	15	-	-	-	36	16	62	51	69	278	86	5.1	1460	0	-	-	-	48	36	30	54	366	429	7.6	440	960	-	-	-	30	17	12	22	770	933	7.4	270	3220	-	-	-	-
259	770	15	46	18.5	1.9	36	17	60	49	69	293	80	5.3	1530	0	1.2e9	61	48	49	42	34	60	374	397	7.5	540	970	2.2e6	68	0	31	13	9	19	778	938	7.3	160	3350	-	-	-	101
260	780	15	45	18.5	1.9	36	17	48	39	81	231	81	5.7	1010	0	-	-	-	49	40	33	51	372	373	7.5	550	830	-	-	-	31	11	7	20	774	898	7.3	210	3340	-	-	-	-
261	770	15	46	18.6	1.8	38	16	41	35	60	284	172	4.9	1550	0	-	-	-	48	46	37	56	374	399	7.1	520	980	-	-	-	31	22	15	25	705	844	7.2	280	2800	-	-	-	-
262	760	16	-	-	-	33	15	44	36	63	267	144	4.8	1440	0	-	-	-	48	41	33	50	361	286	7.3	540	820	-	-	-	31	12	8	11	745	883	7.3	290	3060	-	-	-	-
263	760	16	-	-	-	36	17	39	33	56	266	145	4.8	1530	0	-	-	-	49	41	33	46	376	380	7.2	570	810	-	-	-	31	30	20	28	773	953	7.0	260	3200	-	-	-	-
264	760	16	46	18.5	1.9	36	16	-	-	-	-	-	-	-	-	-	-	-	49	-	-	-	-	-	-	-	-	-	-	-	31	-	-	-	-	-	-	-	-	-	-	-	-
265	760	16	-	-	-	36	16	55	44	65	285	138	5.6	1200	0	6.0e9	99	62	48	42	34	54	372	399	7.4	500	820	7.0e5	67	0	31	40	27	46	762	999	7.3	330	3220	1.5e5	447	10	-
266	750	15	46	18.5	2.1	37	16	51	42	65	293	122	5.6	1190	130	-	-	-	49	42	34	49	377	405	7.6	490	900	-	-	-	32	40	27	46	776	963	7.2	350	3190	-	-	-	153
Steady State Period 7 Ends																			Average Values for the Last 10 days of Period																								
-	765	15	46	18.5	1.9	36	16	49	40	66	268	116	5.2	1337	14	-	-	-	48	41	33	53	372	386	7.4	530	891	-	-	-	31	22	14	25	761	924	7.3	258	3190	-	-	-	-
267	730	14	44	18.6	1.8	37	15	60	48	76	253	102	5.7	930	30	-	-	-	48	36	29	62	393	419	7.7	500	1080	-	-	-	32	31	21	16	778	918	7.4	170	3310	-	-	-	-
268	720	15	44	18.6	1.8	36	14	47	36	62	292	120	5.3	1480	0	-	-	-	47	44	35	56	408	473	7.4	660	970	-	-	-	32	29	19	40	772	999	7.4	300	3250	-	-	-	-
269	730	14	-	-	-	36	14	64	52	76	313	115	5.2	1710	0	-	-	-	46	46	36	53	416	443	7.1	680	970	-	-	-	31	21	14	31	758	933	7.3	240	3200	-	-	-	-
270	730	15	-	-	-	36	14	45	38	55	249	80	5.4	1070	0	-	-	-	45	44	35	59	412	458	7.2	760	840	-	-	-	30	26	18	26	728	932	7.3	440	2960	-	-	-	-
271	720	15	41	18.7	1.7	38	15	40	33	59	275	124	5.6	1180	0	-	-	-	46	43	35	61	410	469	7.3	870	720	-	-	-	29	15	10	26	760	884	7.3	230	3240	-	-	-	-
272	710	14	-	-	-	36	15	42	34	51	287	108	5.9	1160	0	-	-	-	45	44	36	64	404	485	7.3	580	1160	-	-	-	29	23	16	37	739	907	7.4	250	3250	-	-	-	-
273	690	15	40	18.6	1.9	37	13	59	47	90	254	103	5.6	1180	0	-	81	32	44	42	34	70	393	456	7.7	500	1150	-	272	5	28	16	11	29	740	858	7.4	140	3230	-	-	-	396
274	700	16	40	18.8	1.7	36	13	60	47	84	222	98	5.7	880	0	-	-	-	43	38	30	50	391	471	7.7	730	1000	-	-	-	27	18	13	28	714	838	7.4	240	3090	-	-	-	-
275	690	17	39	18.9	1.5	36	12	56	43	86	358	150	5.5	1910	0	-	-	-	42	46	37	71	432	523	7.0	900	800	-	-	-	27	13	9	14	706	838	7.6	220	3010	-	-	-	-
276	640	16	-	-	-	36	14	56	44	71	290	96	5.4	1450	0	-	-	-	42	47	37	61	423	503	7.1	830	820	-	-	-	27	10	7	17	703	818	7.2	180	3070	-	-	-	-

Table 2a.9 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase I Oxygenation with Air

Day	Aeration					Feed Sludge												Aerobic Sludge												Anaerobic Sludge											
No	Q_{AIR}	T_{in}	T_{out}	O_2	CO_2	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	T_{fd}	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	ASC	VI	SRF
	m^3/h	°C	°C	%	%	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	°C	g/l	g/l	g/l	mS/m	mg/l	–	mg/l	mg/l	/cl	/g	%	m/kg
277	620	14	-	-	-	37	12	60	49	71	268	76	5.2	1380	0	-	-	-	41	40	32	54	400	463	7.2	730	920	-	-	-	26	7	5	15	671	782	7.3	150	3030	-	-	-	-
278	750	16	40	18.9	1.6	36	13	-	-	-	-	-	-	-	-	-	-	-	42	38	30	61	381	419	7.2	700	750	-	-	-	25	9	6	24	677	783	7.3	240	2920	-	-	-	-
279	760	17	-	-	-	36	12	62	50	90	369	197	5.4	1870	0	-	-	-	44	43	35	56	390	439	7.2	610	900	-	-	-	25	19	13	22	664	778	7.2	150	2820	-	-	-	-
280	750	16	42	18.7	1.7	50	13	66	54	70	333	137	5.3	1720	0	-	-	-	45	47	38	60	386	443	7.1	860	560	-	-	-	25	22	13	16	650	728	7.1	140	2710	-	-	-	-
281	770	16	42	18.6	1.9	36	15	88	74	95	236	86	5.7	870	0	-	-	-	45	37	32	75	438	533	7.0	1280	450	-	-	-	25	19	14	22	628	698	7.1	210	2670	-	-	-	-
282	770	16	45	18.6	1.8	36	13	52	42	90	312	134	5.4	1450	0	-	-	-	46	47	38	68	448	559	7.4	950	900	-	-	-	25	25	17	28	638	718	7.1	130	2790	-	-	-	-
283	770	17	-	-	-	29	12	50	40	71	284	120	5.3	1270	0	-	-	-	46	41	33	61	463	575	7.3	940	890	-	-	-	24	26	17	26	642	708	7.2	120	2730	-	-	-	-
284	770	16	-	-	-	36	14	52	43	68	212	84	5.5	770	0	-	-	-	46	48	38	61	443	541	7.4	840	1080	-	-	-	24	25	17	37	629	753	7.0	130	2740	-	-	-	-
Steady State Period 8 Begins																																										
285	770	17	-	-	-	28	16	51	43	94	260	124	5.4	1190	0	3.3e10	-	-	48	45	36	60	395	483	7.4	820	830	3.4e5	-	-	24	20	14	23	640	793	7.2	310	2660	-	-	-	-
286	750	15	-	-	-	36	16	52	43	83	221	84	5.7	660	0	-	-	-	49	43	36	53	381	437	7.3	580	900	-	-	-	24	16	11	18	646	718	7.2	110	2810	-	-	-	-
287	780	15	46	18.5	1.9	37	17	48	38	85	276	128	5.4	1310	0	-	-	-	49	49	35	79	360	378	7.4	550	920	-	-	-	24	42	29	30	661	748	7.2	180	2790	-	-	-	-
288	770	17	47	18.4	2.0	36	16	51	40	95	219	69	5.5	800	0	-	-	-	50	43	35	42	312	315	7.4	460	820	-	-	-	24	45	32	48	656	783	7.1	170	2860	-	-	-	-
289	750	17	47	18.5	1.9	37	16	43	35	60	255	87	5.2	1300	0	-	-	-	50	-	-	-	-	-	-	-	-	-	-	-	24	10	7	10	561	615	7.2	200	2760	-	-	-	-
290	770	17	-	-	-	39	16	55	44	73	258	106	5.3	1290	0	-	-	-	49	38	31	42	270	251	6.9	470	530	-	-	-	24	13	9	11	570	638	7.1	200	2510	-	-	-	-
291	770	18	-	-	-	37	17	57	47	95	244	112	5.3	1180	0	-	-	-	49	48	38	76	261	227	6.9	500	440	-	-	-	24	10	7	9	575	633	7.1	160	2490	-	-	-	-
292	770	17	47	18.6	1.7	36	16	46	39	93	331	146	5.0	2370	0	-	-	-	48	48	40	62	282	251	6.6	820	280	-	-	-	24	-	-	-	-	-	-	-	-	-	-	-	-
293	750	17	-	-	-	33	17	54	44	74	330	121	5.2	2000	0	-	-	-	49	39	31	55	265	238	7.0	710	290	-	-	-	24	12	8	12	558	643	7.2	380	2160	-	-	-	-
294	750	20	51	18.1	2.2	37	18	60	49	-	318	168	5.3	1740	0	-	-	-	51	44	36	-	297	287	7.2	620	610	-	-	-	25	20	14	-	574	663	7.0	330	2320	-	-	-	-
295	760	17	53	18.0	2.5	36	18	61	49	-	291	130	5.5	1360	0	-	-	-	52	46	37	-	323	347	7.6	520	780	-	-	-	25	18	12	-	566	628	7.0	240	2320	-	-	-	-
296	750	17	52	17.8	2.4	35	19	55	44	80	323	171	5.3	2180	0	-	-	-	53	47	37	55	367	409	7.4	680	900	-	-	-	25	14	9	18	551	679	7.4	270	2400	-	-	-	-
297	760	17	-	-	-	36	20	61	49	77	258	118	5.5	1440	0	-	-	-	53	47	37	57	380	445	7.4	690	1010	-	-	-	25	17	12	14	570	698	7.2	220	2530	-	-	-	-
298	780	14	-	-	-	36	19	59	48	80	260	120	5.4	1540	0	-	-	-	51	46	36	56	392	476	7.8	720	960	-	-	-	25	22	15	24	578	708	7.2	280	2460	-	-	-	-
299	780	15	48	18.5	1.9	36	16	56	45	90	260	123	5.3	1350	0	-	-	-	50	47	36	60	407	467	7.4	590	1110	-	-	-	26	20	14	28	604	708	7.1	160	2590	-	-	-	-
300	770	16	-	-	-	38	16	49	38	99	429	191	5.8	1050	0	-	-	-	46	46	36	56	429	501	7.3	650	1100	-	-	-	25	23	16	30	602	698	7.2	140	2660	-	-	-	-
301	780	15	44	18.6	1.7	36	17	56	43	95	416	114	5.5	1460	0	-	-	-	47	47	35	49	416	440	7.5	420	1340	-	-	-	26	20	14	22	622	700	7.4	220	2660	-	-	-	-
302	770	15	46	18.5	1.9	36	18	55	42	85	360	80	5.7	1020	0	-	-	-	49	43	34	53	360	410	7.8	490	1010	-	-	-	26	20	14	34	627	700	7.3	190	2710	-	-	-	-
303	770	15	-	-	-	37	20	55	41	100	358	175	5.4	1870	0	-	-	-	51	-	-	-	-	-	-	-	-	-	-	-	26	-	-	-	-	-	-	-	-	-	-	-	-
304	780	17	-	-	-	37	21	53	42	73	379	148	5.5	1700	0	5.0e9	-	-	52	46	36	48	379	399	7.5	500	1020	1.0e6	-	-	26	20	14	22	630	708	7.3	280	2650	-	-	-	-
305	780	17	-	-	-	36	20	54	42	85	301	130	5.4	1380	0	-	-	-	51	46	36	58	381	429	7.6	540	990	-	-	-	26	-	-	-	-	-	-	-	-	-	-	-	190
306	770	15	50	18.4	2.1	36	21	42	33	73	242	129	5.6	1160	0	-	-	-	52	44	34	59	386	399	7.8	520	930	-	-	-	26	20	14	23	644	715	7.4	240	2660	-	-	-	-
307	760	16	-	-	-	37	20	49	40	81	260	128	5.6	1100	0	-	-	-	51	45	35	66	402	449	7.6	580	1020	-	-	-	26	20	14	23	641	748	7.4	140	2810	-	-	-	-
308	760	14	46	18.5	1.9	39	18	38	30	56	254	109	5.6	1140	0	-	-	-	49	44	34	65	395	445	7.6	590	960	-	-	-	26	-	-	-	-	-	-	-	-	-	-	-	-
309	770	16	45	18.5	1.7	37	18	49	39	73	241	96	5.6	1030	0	-	-	-	48	43	33	63	392	421	7.4	530	1000	-	-	-	26	21	14	-	660	748	7.4	300	2730	-	-	-	100
310	770	15	-	-	-	36	19	45	35	-	304	186	5.4	1300	0	-	-	-	48	41	32	-	348	356	7.6	360	890	-	-	-	26	20	14	-	634	653	7.5	140	2560	-	-	-	-
311	770	15	-	-	-	36	22	36	28	-	252	153	5.4	1150	0	-	-	-	49	-	-	-	-	-	-	-	-	-	-	-	26	21	14	-	661	785	7.3	170	2680	-	-	-	-
312	760	16	-	-	-	36	20	48	38	-	247	135	5.4	1130	0	-	-	-	49	41	32	-	370	415	7.5	390	960	-	-	-	26	21	14	-	660	773	7.2	160	2710	-	-	-	-
Steady State Period 8 Ends																		Average Values for the Last 10 days of Period																							
-	769	16	47	18.5	1.9	37	20	47	37	77	284	139	5.5	1296	0	-	-	-	50	44	34	60	382	414	7.6	501	971	-	-	-	26	20	14	23	647	733	7.4	204	2686	-	-	-	-

APPENDIX 3

CHEMICAL DATA RELATING TO THE LABORATORY SCALE STUDY

3a.1 Objective of the Study

Appendix 3 contains a complete compilation of all the chemical data relating to the laboratory scale study, undertaken to determine whether or not aerobic conditioning is able to reduce the retention time in the anaerobic digester below that normally associated with conventional anaerobic digestion (15days).

3a.2 The Study Period

Two laboratory scale digesters (Digester 1 and Digester 2), each with an operating capacity of 6ℓ, were operated for a period of 50 days, at a sludge retention time of 8 days. The study commenced on the 12th June 1990 (designated day 1) and was terminated on the 31st July 1990 (designated day 50). The commencement and termination days of this study are days 251 and 300 of the dual digester phase I investigation period.

3a.3 Monitoring Aspects

Digester 1 was fed with raw primary sludge taken from the gravity thickener daily composite sample, whilst Digester 2 was fed with sludge taken from the aerobic reactor daily composite sample. The results of the analyses on these samples are given in Appendix 2 (days 251 to 300) and are repeated here for ease of comparison.

Samples of effluent sludge from Digester 1 and Digester 2 were taken three times weekly and tested for TS, VS, NH_4^+, Conductivity, Bicarbonate and Volatile Acid Alkalinity concentrations. Gas composition and production were not measured, and digester stability or failure was assessed on the results of the measured sludge parameters. The study on Digester 1 was terminated on day 28 after problems were experienced with the desludging pump ; by this stage the pH had dropped below 6.0, indicating process failure.

3a.4 Compilation of the Results

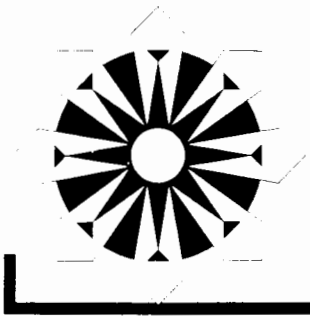
A full description of the units and the mode of operation is given in Section 2.5. The results of the study are summarised and discussed in Section 4.5. All the results from the laboratory scale study are listed in the Tables 3a.1 and 3a.2 below:

Table 3a.1 Chemical Data for the Laboratory Scale Digester No. 1

	Feed Sludge to Digester 1 (ex gravity thickener)							Effluent Sludge from Digester1						
Day No.	TS g/l	VS g/l	NH_4 mg/l	COND mS/m	pH –	VAA mg/l	BA mg/l	TS g/l	VS g/l	NH_4 mg/l	COND mS/m	pH –	VAA mg/l	BA mg/l
1	65.8	51.7	91	260	5.4	1370	0	13.7	9.2	923	739	7.5	260	3060
2	57.4	46.1	64	274	5.0	1810	0	-	-	-	-	-	-	-
3	62.8	50.5	59	225	5.2	1260	0	-	-	-	-	-	-	-
4	45.7	37.4	97	261	5.1	1710	0	-	-	-	-	-	-	-
5	50.9	41.0	82	224	5.2	1330	0	-	-	-	-	-	-	-
6	57.0	46.9	62	202	5.2	1270	0	-	-	-	-	-	-	-
7	41.4	34.8	73	218	5.2	1120	0	-	-	-	-	-	-	-
8	62.5	51.4	86	278	5.1	1460	0	22.0	16.0	683	612	7.0	679	1871
9	59.6	49.2	80	293	5.3	1530	0	-	-	-	-	-	-	-
10	48.3	39.2	81	231	5.7	1010	0	-	-	-	-	-	-	-
11	40.6	34.7	172	284	4.9	1550	0	25.4	19.6	838	650	6.8	2070	882
12	43.6	36.5	144	267	4.8	1440	0	-	-	-	-	-	-	-
13	39.3	32.8	145	266	4.8	1530	0	-	-	-	-	-	-	-
14	-	-	-	-	-	-	-	-	-	-	-	-	-	-
15	54.7	44.2	138	285	5.6	1200	0	24.7	19.1	718	650	6.3	1661	659
16	50.7	42.0	122	293	5.6	1190	130	-	-	-	-	-	-	-
17	60.0	48.2	102	253	5.7	930	30	30.9	20.3	798	658	6.9	1647	953
18	46.6	36.1	120	292	5.3	1480	0	-	-	-	-	-	-	-
19	64.0	51.7	115	313	5.2	1710	0	-	-	-	-	-	-	-
20	44.8	37.6	80	249	5.4	1070	0	-	-	-	-	-	-	-
21	40.0	33.0	124	275	5.6	1180	0	32.2	21.3	803	652	6.5	1860	471
22	42.5	33.9	108	287	5.9	1160	0	-	-	-	-	-	-	-
23	58.8	46.8	103	254	5.6	1180	0	-	-	-	-	-	-	-
24	60.5	47.0	98	222	5.7	880	0	34.4	27.6	758	619	6.5	2350	0
25	56.4	43.1	150	358	5.5	1910	0	-	-	-	-	-	-	-
26	55.8	44.2	96	290	5.4	1450	0	-	-	-	-	-	-	-
27	59.5	49.4	76	268	5.2	1380	0	-	-	-	-	-	-	-
28	-	-	-	-	-	-	-	40.1	32.0	610	605	5.9	2100	0
29	62.4	49.7	197	369	5.4	1870	0	-	-	-	-	-	-	-
30	66.2	54.2	137	333	5.3	1720	0	-	-	-	-	-	-	-
31	88.2	73.8	86	236	5.7	870	0	-	-	-	-	-	-	-
32	52.4	42.2	134	312	5.4	1450	0	-	-	-	-	-	-	-
33	49.7	39.9	120	284	5.3	1270	0	-	-	-	-	-	-	-
34	51.5	42.8	84	212	5.5	770	0	-	-	-	-	-	-	-
35	50.8	42.9	124	260	5.4	1190	0	-	-	-	-	-	-	-
36	52.5	43.3	84	221	5.7	660	0	-	-	-	-	-	-	-
37	48.1	38.2	128	276	5.4	1310	0	-	-	-	-	-	-	-
38	51.3	40.1	69	219	5.5	800	0	-	-	-	-	-	-	-
39	-	-	-	-	-	-	-	-	-	-	-	-	-	-
40	54.9	44.3	106	258	5.3	1290	0	-	-	-	-	-	-	-
41	57.4	47.0	112	244	5.3	1180	0	-	-	-	-	-	-	-
42	46.4	38.8	146	331	5.0	2370	0	-	-	-	-	-	-	-
43	54.3	44.5	121	330	5.2	2000	0	-	-	-	-	-	-	-
44	60.0	48.9	168	318	5.3	1740	0	-	-	-	-	-	-	-
45	61.3	49.2	130	291	5.5	1360	0	-	-	-	-	-	-	-
46	55.3	44.2	171	323	5.3	2180	0	-	-	-	-	-	-	-
47	60.9	48.9	118	258	5.5	1440	0	-	-	-	-	-	-	-
48	58.6	47.5	120	260	5.4	1540	0	-	-	-	-	-	-	-
49	56.0	45.4	123	260	5.3	1350	0	-	-	-	-	-	-	-
50	49.1	38.5	191	429	5.8	1050	0	-	-	-	-	-	-	-

Table 3a.2 Chemical Data for the Laboratory Scale Digester No. 2

	Feed Sludge to Digester 2 (ex aerobic reactor)							Effluent Sludge from Digester2						
Day No.	TS g/l	VS g/l	NH_4 mg/l	COND mS/m	pH –	VAA mg/l	BA mg/l	TS g/l	VS g/l	NH_4 mg/l	COND mS/m	pH –	VAA mg/l	BA mg/l
1	37.3	30.4	541	434	7.7	790	970	13.7	9.2	923	739	7.5	260	3060
2	41.2	33.0	572	464	7.8	870	980	-	-	-	-	-	-	-
3	47.2	37.3	569	461	7.8	700	1050	-	-	-	-	-	-	-
4	45.2	36.2	551	448	7.5	710	1090	-	-	-	-	-	-	-
5	44.4	35.6	479	406	7.6	550	1080	-	-	-	-	-	-	-
6	44.6	35.7	445	404	7.4	670	970	-	-	-	-	-	-	-
7	40.1	32.7	402	379	7.6	620	930	-	-	-	-	-	-	-
8	36.0	29.6	429	366	7.6	440	960	19.0	14.6	823	683	7.1	734	2361
9	42.0	34.1	397	374	7.5	540	970	-	-	-	-	-	-	-
10	40.3	32.8	373	372	7.5	550	830	-	-	-	-	-	-	-
11	46.5	37.4	399	374	7.1	520	980	19.7	15.3	853	733	7.2	876	2399
12	41.2	33.1	286	361	7.3	540	820	-	-	-	-	-	-	-
13	40.8	32.9	380	376	7.2	570	810	-	-	-	-	-	-	-
14	-	-	-	-	-	-	-	-	-	-	-	-	-	-
15	41.8	34.0	399	372	7.4	500	820	19.0	14.4	839	721	6.9	1062	1888
16	41.8	34.0	405	377	7.6	490	900	-	-	-	-	-	-	-
17	35.5	28.8	419	393	7.7	500	1080	19.7	15.4	849	704	7.1	955	1948
18	44.4	35.4	473	408	7.4	660	970	-	-	-	-	-	-	-
19	46.1	36.5	443	416	7.1	680	970	-	-	-	-	-	-	-
20	44.0	35.1	458	412	7.2	760	840	-	-	-	-	-	-	-
21	43.3	34.7	469	410	7.3	870	720	20.6	15.5	859	711	7.0	1045	1970
22	44.3	35.7	485	404	7.3	580	1160	-	-	-	-	-	-	-
23	42.5	33.6	456	393	7.7	500	1150	-	-	-	-	-	-	-
24	38.2	30.2	471	391	7.7	730	1000	20.9	15.7	862	705	7.1	1050	1960
25	46.1	36.7	523	432	7.0	900	800	-	-	-	-	-	-	-
26	46.9	37.0	503	423	7.1	830	820	-	-	-	-	-	-	-
27	40.2	31.9	463	400	7.2	730	920	-	-	-	-	-	-	-
28	-	-	-	-	-	-	-	22.0	16.7	862	721	7.0	1010	2040
29	43.1	34.9	439	390	7.2	610	900	-	-	-	-	-	-	-
30	47.4	38.5	443	386	7.1	860	560	-	-	-	-	-	-	-
31	36.6	31.9	533	438	7.0	1280	450	-	-	-	-	-	-	-
32	47.4	38.2	559	448	7.4	950	900	24.3	18.2	841	702	7.1	925	2120
33	41.1	33.3	575	463	7.3	940	890	-	-	-	-	-	-	-
34	47.5	38.3	541	443	7.4	840	1080	-	-	-	-	-	-	-
35	44.9	36.3	483	395	7.4	820	830	-	-	-	-	-	-	-
36	43.4	35.5	437	381	7.3	580	900	25.5	18.8	872	730	7.1	950	2060
37	48.8	35.3	378	360	7.4	550	920	-	-	-	-	-	-	-
38	42.7	34.6	315	312	7.4	460	820	-	-	-	-	-	-	-
39	-	-	-	-	-	-	-	25.0	18.6	813	697	7.0	1065	1935
40	38.2	30.9	251	270	6.9	470	530	-	-	-	-	-	-	-
41	47.8	38.5	227	261	6.9	500	440	-	-	-	-	-	-	-
42	47.8	40.3	251	282	6.6	820	280	24.9	19.0	849	708	7.1	950	2140
43	38.7	30.6	238	265	7.0	710	290	-	-	-	-	-	-	-
44	44.4	35.7	287	297	7.2	620	610	-	-	-	-	-	-	-
45	46.1	36.6	347	323	7.6	520	780	-	-	-	-	-	-	-
46	47.0	37.4	409	367	7.4	680	900	25.2	18.9	881	737	7.1	945	2255
47	46.9	37.0	445	380	7.4	690	1010	-	-	-	-	-	-	-
48	45.7	36.0	476	392	7.8	720	960	-	-	-	-	-	-	-
49	47.1	36.4	467	407	7.4	590	1110	-	-	-	-	-	-	-
50	46.4	36.3	501	429	7.3	650	1100	25.0	18.7	879	730	7.1	925	2225



APPENDIX 4

OXYGEN TRANSFER STUDIES: AIR OXYGENATION DEVICE

4a.1 Introduction

Prior to the start of phase I, a non-steady state aeration test of de-oxygenated tap water (Bratby, 1977) was performed on the aerobic reactor of the dual digester. The objective of the test was to determine the oxygen transfer efficiency ***OTE*** and the maximum oxygen transfer rate OTR^{max} of the course bubble aeration equipment in the aerobic reactor.

4a.2 The Non-Steady State Aeration Test: Experimental Data

Tap water in the reactor (184 m^3) was de-oxygenated with sodium sulphite (150 mg/l) and cobalt chloride was added (2 mg/l) to catalyse the reaction. In conformity with expected normal operation (oxygenation with air alone), one liquid ring compressor and one mixing pump were in operation for the test; the former was set at its maximum dry air flow rate of 780 m^3(STP)/h (a continuously operating maximum of 760 m^3(STP)/h was set in the investigation) while the latter operated at a predicted flow rate of 1000 m^3/h giving a turn over time of about 11 minutes (obtained from estimated system head loss and pump characteristic curves). Samples of the aerated water were taken at 30 second intervals on the suction side of the mixing pump. Dissolved oxygen levels were subsequently measured using the Winkler method (Standard Methods, 1989). The ambient temperature at the time of the test was 20°C, the atmospheric pressure was measured at 771mmHg, and the temperature of the water in the reactor was 17°C. At the end of the test, the measured and calculated saturation dissolved oxygen concentration of the water in the reactor were in agreement at 9.8mg/l. The power drawn by the liquid ring compressor was 20kW and by the mixing pump was 10kW. The dissolved oxygen data obtained during the test are contained Table 4a.1 below:

Table 4a.1 Dissolved Oxygen Data for the Non-Steady State Aeration Test

t (s)	C_t (mgO/l)	$(C_s - C_t)$	$\ln(C_s - C_t)$	t (s)	C_t (mgO/l)	$(C_s - C_t)$	$\ln(C_s - C_t)$
0	0.1	9.7	2.27	270	7.0	2.8	1.03
30	1.5	8.3	2.12	300	7.4	2.4	0.88
60	2.6	7.2	1.97	330	7.8	2.0	0.69
90	3.4	6.4	1.86	360	8.1	1.7	0.53
120	4.2	5.6	1.72	390	8.4	1.4	0.34
150	4.8	5.0	1.61	450	9.0	0.8	-
180	5.7	4.1	1.41	510	9.6	0.2	-
210	6.2	3.6	1.28	570	9.8	0.0	-
240	6.7	3.1	1.13	630	9.8	0.0	-

4a.3 Determination of the Mass Transfer Coefficient for the Aeration Device

The rate of oxygenation of a body of water by a particular device is given by the following equation (Lewis and Whitman, 1924):

$$\frac{dC}{dt} = K_{La} \cdot (C_s - C) \quad \text{...mg}(O_2)/l \quad (4a.1)$$

where:

C = concentration of dissolved oxygen in the water (mg(O_2)/l)

C_s = saturation concentration of oxygen in equilibrium with the gas phase (mg(O_2)/l)

K_{La} = mass transfer coefficient for the aeration device (s^{-1})

Integration of Eq 4a.1 yields:

$$\ln(C_s - C_t) = K_{La} \cdot t + \ln(C_s - C_0) \quad \text{...} \quad (4a.2)$$

where:

C_t = concentration of dissolved oxygen at time t (mg(O_2)/l)

C_0 = concentration of dissolved oxygen at time 0 (mg(O_2)/l)

By plotting $\ln(C_s - C_t)$ as ordinate versus t as abscissa, a straight line is obtained in the central portion of the curve with slope equal to K_{La} (see Fig. 4a.1).

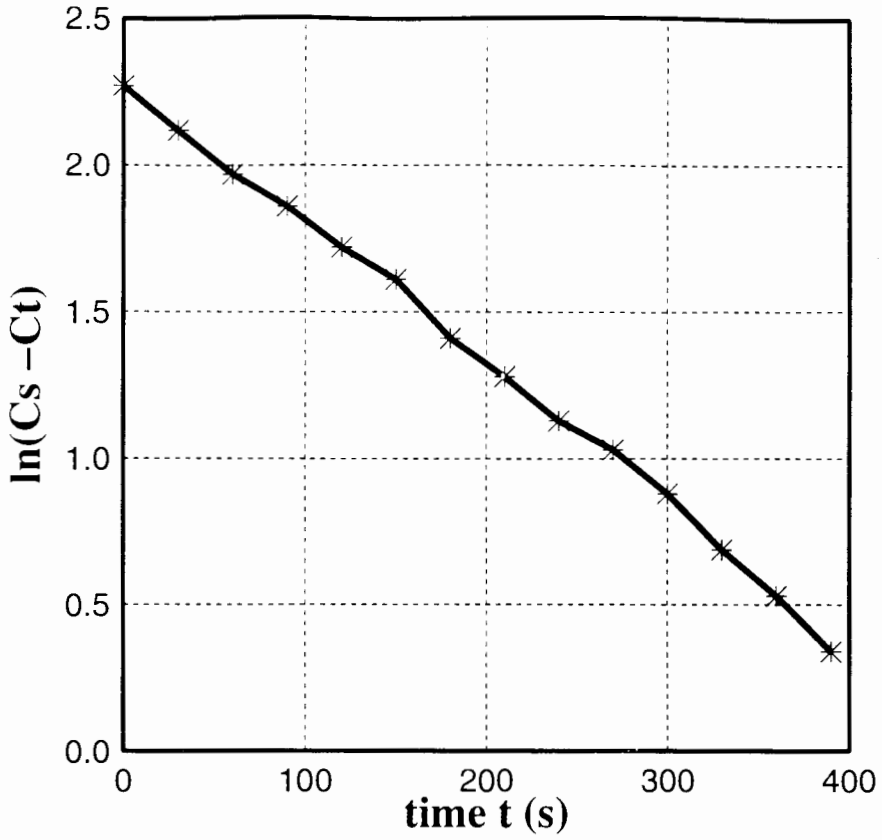


Figure 4a.1 The Non-Steady State Aeration Test: Plot of $\ln(C_s - C_t)$ versus t

Linear regression of the appropriate data in Table 4a.1 between **t**=90s and **t**=300s yields:

$$\ln(C_s - C_t) = -4.72 \times 10^{-3}.t + 2.28 \quad \ldots \quad (4a.3)$$

Hence, at 17°C (the temperature of the water in the reactor) the mass transfer coefficient K_{La} is equal to 4.72x 10^{-3} s^{-1} (17.0 h^{-1}). K_{La} however, increases with increased temperature due to the increased rate of diffusion of oxygen from the gas to the liquid phase. The temperature effect on K_{La} has been predicted by the following relationship (Eckenfelder and Ford, 1968):

$$K_{La}^{T} = K_{La}^{20°C}.\theta^{(T-20°)} \quad \ldots s^{-1} \quad (4a.4)$$

Values for the constant θ for diffused aeration systems are normally within the range 1.02 to 1.028. Accepting a value of 1.024, the mass transfer coefficient at STP is as follows

The Mass Transfer Coefficient at (STP) $K_{La}^{20°C}$ = 5.07x 10^{-3} s^{-1} (18.3 h^{-1})

4a.4 The Oxygenation Capacity

The oxygenation capacity **OC** of an aeration device is defined as the mass of oxygen that the device can introduce into a completely de-oxygenated body of water per unit time under standard conditions ie at an atmospheric pressure of 760 mmHg and temperature of 20°C using clean tap water.

For an aerator started up in a completely de-oxygenated body of water the rate of increase in concentration of dissolved oxygen at time **t** and the corresponding dissolved oxygen concentration C_t are both given by Eq. 4a.1 above. At start-up $C_t = 0$ and $dC_t/dt = K_{La}.C_s$. It follows therefore that, the oxygenation capacity **OC** is equal to the rate of increase in dissolved oxygen concentration when $C_t = 0$ multiplied by the volume of water, ie

$$\mathbf{OC} = \frac{dC}{dt}.V = K_{La}.C_s.V \qquad \text{...mg}(O_2)/\text{s} \quad (4a.5)$$

From the non-steady state aeration test $K_{La}^{20°C} = 5.07 \times 10^{-3}\ s^{-1}$, the saturation concentration of dissolved oxygen C_s at (STP) = 9.07 mg(O_2)/l, and the volume **V** of tap water in the reactor = 184×10^3 l. Application of Eq. 4a.5 yields:

$$\mathbf{OC} = 5.07 \times 10^{-3} \times 9.07 \times 184 \times 10^3 = 8460 \qquad \text{...mg}(O_2)/\text{s} \quad (4a.6)$$

Converting the above result into units of kg(O_2)/h the oxygenation capacity is as follows

The Oxygenation Capacity OC = 30.5kg(O_2)/h

4a.5 The Maximum Oxygen Transfer Rate

The oxygenation capacity **OC** was defined above as the mass of oxygen that the device can introduce into a completely de-oxygenated body of water per unit time under standard conditions. In Section 3.1.7 OTR^{max} was defined as the volume specific maximum oxygen transfer rate which could be generated by the aeration system (kg$(O_2)/m^3$.h). The two terms are linked via the process volume **V** viz:

$$OTR^{max} = \frac{\mathbf{OC}}{V} = \frac{30.5}{184} \quad (@\ 20°C) \qquad \text{...kg}(O_2)/m^3.\text{h} \quad (4a.7)$$

The Maximum Oxygen Transfer Rate OTR^{max} = 0.167kg$(O_2)/m^3$.h

4a.6 The Oxygen Transfer Efficiency

The oxygen supply rate ***OSR*** was defined in Section 3.1.7 as the mass flow rate of oxygen supplied to the reactor per unit process volume ($kg(O_2)/m^3.h$). In terms of the volumetric flow rate of dry air into the reactor $Q(AIR)_{in}$ ($m^3(STP)/h$) the oxygen supply rate is given by:

$$OSR = 0.00151 \times Q(AIR)_{in} \quad ...kg(O_2)/m^3.h \quad (3.47)$$

During the aeration test the dry air flow rate was recorded at 780 $m^3(STP)/h$ giving a calculated ***OSR*** of $1.18kg(O_2)/m^3.h$. The oxygen transfer efficiency ***OTE*** was defined in Section 3.1.7 as the percentage mass of oxygen transferred into solution per mass of oxygen supplied to the reactor ie ***OTE*** = ***OTR/OSR*** x 100. It therefore follows that the oxygen transfer efficiency observed during the aeration test is:

$$OTE = \frac{OTR^{max}}{OSR} \times 100 = \frac{0.167}{1.18} \times 100 \quad (@\ 20°C) \quad ...\% \quad (4a.8)$$

The Oxygen Transfer Efficiency *OTE* = 14.0%

4a.7 The Oxygenation Capacity to Power Ratio

The mass of oxygen transferred into solution per unit power consumption is the ratio of the oxygenation capacity **OC** to the power drawn by both the liquid ring compressor (20kW) and the mixing pump (10kW). The mixing pump is included in the calculation as undoubtably the cross cutting mixing action produced by pumping will improve the aeration characteristics of the system. The oxygenation capacity to power ratio is also (confusingly) termed the oxygen transfer rate, to distinguish it from ***OTR*** the symbol $\mathcal{OTR}$ is used. Accordingly,

$$\mathcal{OTR} = \frac{OC}{Power} = \frac{30.5}{30} \quad ...kg(O_2)/kW.h \quad (4a.9)$$

The Oxygenation Capacity to Power Ratio $\mathcal{OTR}$ = 1.0kg(O)/kW.h

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APPENDIX 5

OPERATIONAL, CHEMICAL AND BACTERIOLOGICAL DATA: PHASE II

5a.1 The Evaluation Period: Phase II

Appendix 5 contains a complete tabulation of all the operational, chemical and bacteriological data collected during phase II of the investigation into the dual digestion process at the Athlone Wastewater Treatment Plant. Phase II of the investigation dealt specifically with operation of the aerobic reactor using air plus pure oxygen for oxygenation. Monitoring commenced on the 16th July 1994 (designated day 1 of phase II) and lasted for 22 weeks until the 14th December 1994 (day 152).

5a.2 System Sludge Stream Monitoring

Sludge samples were taken in the same manner as for phase I (see Appendix 2). The feed sludge to the aerobic reactor was drawn from the gravity thickener. An in-line flow meter, placed close to the point of entry to the aerobic reactor, measured the quantity of sludge pumped into the reactor. To take account of the variation of solids concentration in the gravity thickened sludge stream, the plant operators collected a grab sample midway through each feeding session and composited these over 24 hours to make a daily composite sample.

Sludge from the aerobic reactor and anaerobic digester were sampled once every eight hours and composited over 24 hours. The aerobic reactor sludge was sampled at the recirculation mixing pump. The anaerobic digester sludge was sampled at the digester effluent transfer box when sludge was being transferred to the secondary digester.

Samples for bacteriological analyses and dewaterability studies were taken as grab samples on the specific days these analyses were performed.

5a.3 Aerobic Reactor Gas Stream Monitoring

The volumetric influent dry air flow rate to the aerobic reactor $Q(AIR)_{in}$ was monitored by an orifice plate flow meter with the flow recorded on a chart recorder housed in a control room. The flow rate in m^3/h was automatically corrected to STP, 20°C and 760 mmHg as dry air.

The mass flow rate of pure oxygen ($kg(O_2)/h$) to the aerobic reactor $M(O_2)_{in}$ was controlled with a calibrated rotameter (24-120$kg(O_2)/h$) fitted with a needle valve. Readings were taken hourly. Fine adjustments were made when necessary to ensure a constant flow rate. The quantity of oxygen supplied daily from the storage tank was recorded by a mass meter fitted to the tank, providing confirmation of the accuracy of the rotameter.

The oxygen concentration ($\%^{v}/_{v}$) in the aerobic reactor effluent gas stream was monitored approximately three times weekly during each steady state period using a combustion analyser (which also gave the effluent gas temperature). In order to resolve the gas mass balances across the reactor, the oxygen concentration was measured both whilst the air stream was on (denoted as O_2^1 in Tables 5a.1 to 5a.6 below) and whilst the air stream was temporarily switched off (denoted as O_2^2). Unfortunately, the combustion analyser used during phase II was unable to measure the carbon dioxide concentration in the effluent gas stream. Consequently, the value for the respiration quotient Y_{CO2} =0.70 $mol(CO_2)_{gen}/mol(O_2)_{gen}$ determined during phase I was used to resolve the gas mass balance. A full discussion on the aerobic reactor gas mass balance, whilst oxygenation is with air plus pure oxygen, is given in Section 3.3 above.

5a.4 Temperature Measurement

The same eight PT100 resistance thermometers in use during phase I were employed during phase II. The positions of these thermometers was as follows:

- 3 within the aerobic reactor
- 2 in the aerobic reactor recirculation loop $T(SL)_r$
- 1 in the feed sludge line to the aerobic reactor $T(SL)_{in}$
- 1 in the air delivery line to the aerobic reactor $T(AIR)_{in}$
- 1 in the anaerobic digester $T(SL)_d$

The temperatures from the thermometers placed in the recirculation loop were recorded by chart recorder. Guages were used to measure temperatures from the other thermometers with the operating staff manually recording temperatures every hour on specially designed log sheets.

5a.5 Compilation of Process Operational, Chemical and Bacteriological Data

All the operational, chemical and bacteriological data collected during phase II is contained in the following Tables (Table 5a.1 to Table 5a.6 below):

Table 5a.1 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase II Oxygenation with Air + Oxygen

Day	Aeration							Feed Sludge										Aerobic Sludge										Anaerobic Sludge									
No	Q_{AIR}	M_{O2}	T_{in}	T_{out}	O_2^1	O_2^2	I_{mix}	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC
	m^3/h	kg/h	°C	°C	%	%	amp	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl
1	270	24	16	-	-	-	70	26	17	48	40	69	244	93	5.6	1150	30	-	41	29	21	30	175	97	6.8	250	450	-	33	55	31	56	492	594	6.7	730	1800	-
2	270	24	16	-	-	-	70	26	17	54	45	81	261	116	5.4	1410	0	-	44	34	27	35	161	82	6.8	210	410	-	33	63	34	61	500	620	6.8	480	2130	-
3	280	24	15	-	-	-	80	26	17	58	48	87	326	160	5.3	2040	120	-	46	32	25	38	165	93	7.1	130	480	-	33	48	28	52	480	563	6.8	380	2030	-
4	280	24	15	-	-	-	80	28	18	57	46	83	286	122	5.5	1430	100	-	48	31	24	35	197	146	7.3	150	620	-	33	9	7	12	455	486	7.1	268	2090	-
5	280	24	17	-	-	-	75	51	19	67	55	93	351	157	5.2	2280	340	-	50	31	23	38	197	154	7.2	90	670	-	33	30	17	30	460	530	7.1	200	2200	-
Steady State Period 1 Begins																																					
6	280	60	16	-	-	-	80	95	18	59	48	82	263	112	5.5	1330	440	-	50	31	25	34	203	148	7.3	260	600	-	33	18	11	22	456	591	6.9	220	2100	-
7	250	60	15	-	-	-	78	96	17	40	33	61	272	106	5.4	1370	100	-	50	32	25	36	449	350	7.3	140	430	-	33	32	19	37	450	597	7.1	180	2090	-
8	260	60	13	-	-	-	80	95	17	43	32	62	196	86	5.8	740	40	-	50	31	24	36	406	408	7.5	170	620	-	33	36	24	34	483	541	7.0	100	2270	-
9	260	60	14	-	-	-	78	96	16	51	36	60	131	17	6.0	400	80	-	49	35	26	40	458	489	7.3	200	390	-	33	46	25	28	491	563	7.0	90	2390	-
10	260	60	14	45	18.6	24.4	78	96	16	35	30	53	247	73	5.5	1190	160	-	49	33	25	39	436	567	7.3	160	760	-	33	18	14	29	239	597	7.2	190	2660	-
11	240	60	15	-	-	-	75	95	17	50	38	60	222	60	5.8	750	140	-	50	32	25	38	571	689	7.3	260	760	-	33	34	25	28	411	480	7.0	50	2080	-
12	240	60	15	48	18.6	24.3	75	96	18	59	44	65	361	96	5.6	1360	90	-	51	47	30	59	587	641	7.4	430	650	-	33	31	18	36	486	562	7.1	170	2140	-
13	250	60	17	47	18.6	24.2	75	96	18	36	28	50	267	58	5.8	730	80	-	50	38	28	44	575	657	7.3	640	1000	-	35	25	14	20	463	505	7.1	190	2240	-
14	280	60	17	48	18.7	24.3	75	95	20	55	43	75	295	106	5.4	1860	70	-	51	35	26	45	403	651	7.2	370	840	-	36	10	6	11	459	494	7.0	130	2250	-
15	300	60	17	-	-	-	75	96	19	61	48	79	311	112	5.2	2210	270	-	51	38	28	44	495	668	7.3	520	900	-	36	28	17	22	478	524	7.1	80	2410	-
Steady State Period 1 Ends																			Average Values for the Last 7 days of Period																		
-	261	60	16	47	18.6	24.3	76	96	18	50	38	63	262	75	5.6	1210	130	-	50	37	27	44	503	623	7.3	370	760	-	34	27	17	25	432	532	7.1	130	2310	-
Steady State Period 2 Begins																																					
16	610	60	16	-	-	-	75	97	17	44	36	65	343	134	5.4	1460	180	-	51	33	25	36	470	654	7.2	601	1200	-	36	35	23	34	493	563	6.9	120	2380	-
17	600	60	17	48	18.7	21.2	75	96	19	37	29	48	220	112	5.5	980	220	-	52	40	29	44	586	546	7.2	742	1200	-	36	37	24	40	506	591	7.1	120	2450	-
18	610	60	16	-	-	-	75	96	18	41	32	62	220	110	5.5	1010	62	-	53	37	28	42	404	640	7.3	629	1310	-	36	41	25	37	483	591	7.0	190	2900	-
19	610	60	16	-	-	-	75	97	19	50	40	72	292	134	5.4	1760	90	-	53	36	26	40	568	777	7.2	590	1190	-	36	34	22	36	478	555	7.0	260	2350	-
20	600	60	17	51	18.5	21.1	75	96	19	53	43	68	253	120	5.5	1350	100	1.7e10	53	40	31	45	444	515	7.3	950	1240	2.0e6	36	14	9	16	477	569	7.0	180	2350	1.0e4
21	610	60	19	51	18.6	21.0	75	95	20	53	42	70	320	125	5.4	1890	110	-	54	46	36	50	443	543	7.1	880	1380	-	36	28	18	26	495	597	7.1	160	2880	-
22	610	60	19	-	-	-	75	96	21	53	41	55	260	105	6.2	1040	290	-	54	46	34	48	443	484	7.1	950	1220	-	36	19	12	18	533	611	7.0	172	2470	-
23	610	60	17	-	-	-	75	96	18	37	30	44	196	72	5.9	760	80	-	53	44	33	52	448	654	7.1	720	1040	-	36	19	13	19	527	619	7.1	120	2040	-
24	600	60	17	51	18.5	21.1	70	96	19	57	46	75	245	119	5.7	1110	310	-	54	40	29	44	519	647	7.0	980	570	-	37	14	9	14	536	605	7.1	150	2490	-
25	620	60	16	-	-	-	70	97	18	40	31	56	233	94	5.6	1160	90	-	53	45	35	56	451	647	7.2	1220	510	-	37	26	18	26	525	597	7.0	340	2440	-
Steady State Period 2 Ends																			Average Values for the Last 7 days of Period																		
-	609	60	17	51	18.5	21.1	74	96	19	49	39	63	257	110	5.7	1300	150	-	53	43	32	48	474	610	7.1	900	1020	-	36	22	14	22	510	593	7.0	200	2430	-

Table 5a.2 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase II Oxygenation with Air + Oxygen

Day	Aeration							Feed Sludge										Aerobic Sludge										Anaerobic Sludge									
No	Q_{AIR}	M_{O2}	T_{in}	T_{out}	O_2^1	O_2^2	I_{mix}	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC
	m^3/h	kg/h	°C	°C	%	%	amp	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl
Steady State Period 3 Begins																																					
26	610	60	16	-	-	-	95	95	19	56	44	71	245	102	5.5	1280	70	1.0e10	54	37	28	46	412	692	7.1	690	750	6.0e6	37	40	26	37	533	630	7.1	200	2600	8.0e3
27	610	60	16	53	18.8	20.7	98	96	17	41	33	55	199	74	5.6	900	290	-	56	38	30	40	539	554	7.0	530	410	-	37	11	8	12	538	594	7.2	300	2460	-
28	610	60	17	53	18.7	20.9	98	96	17	41	33	53	245	89	5.6	780	100	-	55	41	33	48	463	489	7.1	530	910	-	37	15	10	13	544	627	7.1	220	2670	-
29	610	60	16	-	-	-	95	95	18	62	51	87	230	75	5.2	1230	50	-	56	35	25	49	456	552	7.1	770	810	-	37	12	8	14	534	697	7.3	150	2740	-
30	610	60	16	-	-	-	95	96	18	47	40	61	275	112	5.5	1340	90	-	55	38	30	44	487	450	7.0	330	780	-	37	14	9	14	589	680	7.2	260	2570	-
31	610	60	16	53	18.5	20.7	95	96	19	58	49	89	281	199	5.5	1500	110	8.6e8	56	44	34	40	435	726	7.3	850	1050	3.0e6	37	18	12	11	576	688	7.2	200	2890	6.5e3
32	630	60	17	52	18.6	20.6	96	97	20	49	40	63	248	143	5.5	1220	75	-	56	43	34	54	514	482	7.2	610	1360	-	37	35	22	36	584	748	7.2	90	2970	-
Steady State Period 3 Ends																			Average Values for the Last 7 days of Period																			
-	612	60	16	53	18.7	20.7	96	96	18	51	41	68	246	113	5.5	1180	110	-	55	40	30	46	472	564	7.1	620	870	-	37	21	14	19	557	666	7.2	200	2700	-
Steady State Period 4 Begins																																					
33	620	36	18	-	-	-	96	98	20	48	35	59	247	93	5.4	1220	70	-	53	44	34	48	433	529	7.4	680	1430	-	37	36	24	36	599	745	7.2	150	2940	-
34	0	60	18	-	-	-	96	96	19	50	37	59	197	53	5.7	750	60	-	51	49	35	56	480	563	7.0	820	1110	-	37	33	19	26	609	739	7.1	110	2990	-
35	0	60	18	45	-	53.1	96	94	18	42	32	60	220	78	5.6	990	150	-	50	44	31	43	554	477	7.0	480	1500	-	37	21	13	19	605	748	7.1	190	2810	-
36	0	60	17	-	-	-	95	96	17	44	33	62	202	72	5.8	760	120	-	50	39	28	45	477	490	7.1	620	780	-	37	20	12	18	627	744	7.2	180	3090	-
37	0	60	17	-	-	-	95	96	18	44	35	58	164	60	5.6	550	80	-	51	40	28	47	540	498	7.2	320	720	-	37	15	10	16	626	738	7.2	180	2930	-
38	0	60	20	47	-	51.8	94	94	19	38	31	47	238	97	5.5	1220	100	1.0e9	50	33	25	45	579	439	7.3	360	810	6.0e6	37	24	16	24	619	728	7.1	210	2890	-
39	0	60	18	47	-	54.3	92	96	18	38	31	52	244	90	5.5	1230	460	-	51	38	29	50	586	449	7.4	610	860	-	37	23	15	23	610	703	7.2	210	2860	-
40	0	60	17	46	-	55.4	92	96	18	42	34	53	288	103	5.4	1090	200	-	51	36	27	47	490	444	7.1	500	710	-	38	17	12	16	601	711	7.1	222	2560	-
Steady State Period 4 Ends																			Average Values for the Last 7 days of Period																			
-	0	60	18	46	-	53.7	94	95	18	42	33	56	222	79	5.6	940	170	-	51	40	29	48	529	480	7.2	530	930	-	37	22	14	20	613	571	7.1	180	2880	-
41	600	0	17	-	-	-	-	35	19	-	-	-	-	-	-	-	-	-	48	-	-	-	-	-	-	-	-	-	38	-	-	-	-	-	-	-	-	-
42	370	0	15	-	-	-	-	25	19	-	-	-	-	-	-	-	-	-	46	-	-	-	-	-	-	-	-	-	38	-	-	-	-	-	-	-	-	-
43	370	0	17	-	-	-	-	23	18	-	-	-	-	-	-	-	-	-	44	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
44	360	0	17	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	43	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
45	360	0	17	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	43	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
46	320	0	16	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	42	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
47	320	0	16	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	42	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
48	320	0	19	-	-	-	-	24	20	-	-	-	-	-	-	-	-	-	41	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
49	320	0	17	-	-	-	-	24	17	-	-	-	-	-	-	-	-	-	41	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
50	320	0	18	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	41	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
51	320	0	18	-	-	-	-	24	20	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
52	310	0	21	-	-	-	-	23	23	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
53	320	0	17	-	-	-	-	25	17	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-

Table 5a.3 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase II Oxygenation with Air + Oxygen

Day	Aeration							Feed Sludge										Aerobic Sludge										Anaerobic Sludge									
No	Q_{AIR}	M_{O2}	T_{in}	T_{out}	O_2^1	O_2^2	I_{mix}	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC
	m³/h	kg/h	°C	°C	%	%	amp	m³/d	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl
54	320	0	16	-	-	-	-	24	17	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
55	320	0	18	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
56	340	0	19	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
57	340	0	17	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
58	340	0	17	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
59	350	0	16	-	-	-	-	24	18	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
60	420	0	18	-	-	-	-	24	20	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
61	410	0	19	-	-	-	-	24	21	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
62	340	0	18	-	-	-	-	24	20	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
63	340	0	17	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
64	350	0	17	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
65	350	0	17	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	36	-	-	-	-	-	-	-	-	-
66	340	0	18	-	-	-	-	23	20	-	-	-	-	-	-	-	-	-	39	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
67	440	0	19	-	-	-	-	25	21	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
68	440	0	18	-	-	-	-	25	20	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
69	440	0	17	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
70	440	0	18	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
71	440	0	18	-	-	-	-	24	20	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
72	450	0	19	-	-	-	-	24	20	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
73	450	0	18	-	-	-	-	24	19	-	-	-	-	-	-	-	-	-	40	-	-	-	-	-	-	-	-	-	35	-	-	-	-	-	-	-	-	-
Steady State Period 5 Begins																																					
74	0	24	17	-	-	-	220	76	19	53	43	79	228	85	5.3	1430	150	-	45	45	35	50	370	234	7.0	570	730	-	35	22	15	22	558	712	7.1	170	3160	-
75	0	24	18	-	-	-	205	96	20	47	40	69	206	123	5.5	1270	120	-	51	43	35	52	390	354	7.2	350	1000	-	36	15	11	16	499	662	7.2	190	3420	-
76	0	24	18	50	-	22.8	205	95	20	44	36	62	215	96	5.2	1630	70	-	54	46	35	48	430	510	7.3	364	925	-	36	19	14	20	525	715	7.3	190	3380	-
77	0	24	16	50	-	22.4	205	97	18	59	47	85	289	85	5.5	870	320	-	53	48	32	60	534	489	7.4	510	1310	-	36	22	17	25	558	730	7.2	210	3500	-
78	0	24	16	49	-	21.6	204	96	17	50	40	70	245	97	5.4	1260	230	-	53	50	39	62	610	510	7.5	263	1140	-	37	14	9	12	663	845	7.3	190	3460	-
79	0	24	17	-	-	-	205	96	19	45	38	72	210	120	5.4	1440	150	-	54	45	36	55	543	523	7.4	380	1250	-	37	25	18	27	645	760	7.2	180	3430	-
80	0	24	18	-	-	-	204	96	20	47	38	55	269	154	5.2	1620	90	1.8e9	54	41	32	46	634	490	7.5	290	1090	6.1e6	37	25	18	29	560	812	7.3	200	3470	2.2e4
81	0	24	18	50	-	23.0	205	96	19	47	38	67	301	96	5.6	1190	190	-	55	45	35	49	454	545	7.3	440	870	-	38	20	14	20	645	826	7.3	150	3530	-
82	0	24	18	-	-	-	206	95	19	42	35	64	269	70	5.5	800	250	-	54	43	34	54	499	590	7.5	260	1120	-	38	24	16	24	630	910	7.2	180	3470	-
83	0	24	19	51	-	23.5	205	96	20	56	45	79	289	85	5.3	1530	160	5.8e9	55	43	33	49	556	490	7.4	350	990	3.2e5	39	22	19	29	678	876	7.3	180	3620	4.2e4
Steady State Period 5 Ends																			Average Values for the Last 7 days of Period																		
-	0	24	17	50	-	22.6	205	96	19	49	40	70	267	101	5.4	1240	200	-	54	45	34	54	547	520	7.4	360	1110	-	37	22	16	24	626	823	7.3	180	3500	-

Table 5a.4 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase II Oxygenation with Air + Oxygen

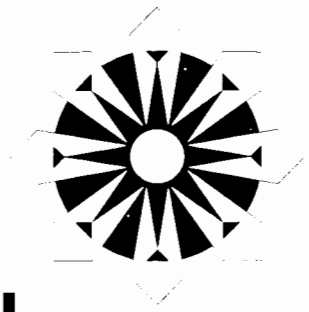
Day	Aeration							Feed Sludge										Aerobic Sludge										Anaerobic Sludge									
No	Q_{AIR}	M_{O2}	T_{in}	T_{out}	O_2^1	O_2^2	I_{max}	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC
	m^3/h	kg/h	°C	°C	%	%	amp	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl
Steady State Period 6 Begins																																					
84	400	24	19	-	-	-	205	96	20	50	42	74	367	132	5.2	1860	70	-	54	46	36	52	489	467	7.5	400	1430	-	39	19	14	20	702	860	7.2	200	3490	-
85	380	24	20	-	18.6	18.7	204	94	20	58	47	82	386	100	5.4	1530	80	-	54	52	40	65	565	485	7.5	150	1120	-	39	23	16	25	689	948	7.3	170	3600	-
86	380	24	17	-	-	-	204	95	19	51	42	71	290	120	5.5	1160	260	-	56	39	30	40	567	552	7.4	220	1260	-	40	25	17	25	664	886	7.3	130	3720	-
87	380	24	17	-	-	-	205	94	19	41	33	59	268	101	5.5	1230	140	-	56	46	36	53	541	545	7.3	190	1160	-	40	22	15	23	704	860	7.4	150	3680	-
88	390	24	19	-	-	-	205	96	19	46	38	65	248	98	5.2	1670	90	2.2e10	56	42	33	50	590	563	7.5	270	1100	1.2e5	41	27	19	27	772	875	7.3	120	3700	8.2e4
89	380	24	21	-	-	-	205	96	20	50	40	70	221	80	5.4	930	270	-	57	46	36	53	675	529	7.3	330	1200	-	41	21	14	21	725	823	7.3	130	3720	-
90	380	24	19	54	18.6	18.8	205	95	19	58	49	79	318	127	5.2	1190	200	-	57	45	34	49	559	652	7.1	270	1200	-	41	16	11	16	707	932	7.2	140	3730	-
91	380	24	21	55	18.7	18.8	210	95	20	48	41	68	296	145	5.3	1970	230	-	57	45	35	46	572	689	7.5	300	1360	-	42	36	21	32	692	949	7.3	180	3710	-
92	370	24	21	55	18.5	18.6	210	95	21	51	43	66	299	138	5.2	1520	280	-	58	52	40	59	643	762	7.2	360	1240	-	42	23	15	23	760	876	7.3	160	3750	-
93	380	24	19	-	-	-	208	96	19	58	49	73	256	74	5.4	1980	80	-	58	51	41	56	598	700	7.2	320	1140	-	43	19	13	21	763	838	7.3	340	3730	-
94	380	24	18	-	-	-	210	95	20	45	36	63	281	88	5.3	1780	380	-	58	39	32	45	550	571	7.2	290	1380	-	43	24	15	23	766	869	7.3	520	3830	-
95	380	24	18	56	18.6	18.7	210	95	19	41	33	59	320	90	5.2	1280	360	-	58	39	31	46	589	568	7.2	120	880	-	44	22	14	24	765	854	7.3	290	3510	-
96	380	24	19	56	18.7	18.9	208	96	21	48	39	56	356	67	5.1	1680	310	6.2e10	59	44	34	50	567	542	7.3	180	880	2.1e4	44	26	18	28	758	845	7.2	390	3520	4.1e3
Steady State Period 6 Ends																			Average Values for the Last 7 days of Period																			
-	379	24	19	55	18.6	18.8	209	95	20	50	41	66	304	104	5.2	1630	260	-	58	45	35	50	583	641	7.2	260	1150	-	43	24	15	24	744	880	7.3	290	3680	-
Steady State Period 7 Begins																																					
97	650	24	18	-	-	-	208	88	18	49	40	53	348	103	5.3	1340	290	-	58	45	36	50	576	638	7.5	300	1500	-	45	22	15	25	740	855	7.2	480	3020	-
98	650	24	18	-	-	-	209	95	19	50	39	68	282	132	5.3	1660	200	-	57	49	39	54	562	596	7.6	310	1230	-	45	28	19	31	764	860	7.2	840	2520	-
99	670	24	18	55	18.6	18.7	208	96	18	49	40	64	264	100	5.2	1810	120	-	57	50	39	56	574	669	7.8	410	1230	-	45	21	15	26	736	766	7.1	920	2108	-
100	680	24	20	-	-	-	209	96	21	52	43	81	239	183	5.4	1800	180	-	58	48	38	57	562	658	7.7	400	1240	-	45	20	14	27	614	738	7.0	940	2270	-
101	690	24	22	-	-	-	205	96	22	56	46	72	273	94	5.4	1900	220	4.0e8	59	42	33	54	618	826	7.9	640	740	<1e3	46	26	18	30	696	645	6.9	830	2360	<1e3
102	690	24	20	-	-	-	205	96	21	63	52	81	230	84	5.3	2210	80	-	59	44	36	57	614	815	7.7	110	880	-	47	28	19	43	612	780	7.0	890	2540	-
103	680	24	21	-	-	-	209	96	21	50	41	57	244	80	5.6	1760	220	-	58	48	38	56	626	787	7.1	540	730	-	47	24	17	27	606	731	6.9	880	2400	-
104	670	24	20	56	18.6	18.6	211	96	22	53	44	75	213	86	5.3	2160	360	8.0e8	58	50	40	56	571	787	7.2	680	780	<1e3	47	34	22	34	613	973	6.9	930	2600	<1e3
105	700	24	21	56	18.6	18.7	211	94	20	51	42	65	263	111	5.5	1950	270	-	59	39	32	46	665	563	7.3	610	850	-	48	37	25	36	710	880	6.9	750	2880	-
106	680	24	21	56	18.7	18.8	205	96	21	53	45	67	228	164	5.3	1800	380	-	59	48	39	62	537	778	7.2	650	1170	-	48	27	19	33	701	987	7.0	640	2740	-
107	640	24	21	-	-	-	205	96	20	54	46	76	247	110	5.1	2160	130	-	58	39	31	46	585	776	7.3	550	910	-	48	33	23	38	725	1050	7.0	410	3560	-
Steady State Period 7 Ends																			Average Values for the Last 7 days of Period																			
-	679	24	21	56	18.6	18.7	207	96	21	54	45	70	242	104	5.4	1990	240	-	59	44	36	54	602	762	7.4	540	860	-	47	30	20	34	666	863	6.9	760	2730	-

Table 5a.5 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase II Oxygenation with Air + Oxygen

Day	Aeration							Feed Sludge										Aerobic Sludge										Anaerobic Sludge									
No	Q_{AIR}	M_{O2}	T_{in}	T_{out}	O_2^1	O_2^2	I_{mix}	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC
	m^3/h	kg/h	°C	°C	%	%	amp	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl
Steady State Period 8 Begins																																					
108	0	36	21	-	-	-	210	96	23	53	43	65	260	180	5.3	1200	140	-	60	43	35	48	554	654	7.2	650	780	-	49	31	22	33	700	967	7.0	440	3870	-
109	0	36	22	-	-	-	208	96	22	51	40	70	206	104	5.4	970	300	6.9e9	61	41	33	52	595	787	7.2	730	980	<1e3	49	30	23	30	747	1110	7.2	420	3640	<1e3
110	0	36	23	57	-	21.8	205	99	24	45	37	58	298	132	5.0	1970	200	-	62	34	27	43	546	658	7.1	300	800	-	50	25	18	29	753	1080	7.3	390	3660	-
111	0	36	20	-	-	-	207	96	21	46	39	62	361	152	5.3	2140	260	-	63	37	30	41	451	806	7.2	950	1060	-	50	32	22	36	790	1010	7.3	350	3560	-
112	0	36	20	57	-	23.0	206	96	21	48	40	63	373	146	5.2	2210	250	-	62	40	33	45	441	739	7.1	1000	890	-	50	20	16	23	789	1010	7.3	450	3570	-
113	0	36	18	-	-	-	207	96	21	46	39	57	366	152	5.2	2090	160	-	62	42	35	52	416	681	7.2	830	860	-	50	23	18	25	785	1060	7.2	340	3500	-
114	0	36	19	-	-	-	206	96	20	40	34	59	269	174	5.2	1320	120	-	62	40	33	49	332	631	7.3	540	720	-	50	28	20	32	758	944	7.3	440	3430	-
115	0	36	19	58	-	24.4	204	96	20	56	48	78	304	179	5.3	1910	150	-	62	54	44	69	362	730	7.4	520	950	-	50	24	17	33	757	951	7.4	410	3530	-
116	0	36	18	57	-	22.1	204	96	19	48	39	64	281	139	5.1	1800	310	2.0e10	61	42	36	48	344	735	7.3	470	1000	<1e3	50	29	21	35	738	993	7.3	480	3520	<1e3
117	0	36	19	-	-	-	205	96	21	50	41	72	266	110	5.3	1550	290	-	61	40	33	44	447	511	7.3	620	990	-	50	29	20	33	734	985	7.5	360	3510	-
Steady State Period 8 Ends																			Average Values for the Last 7 days of Period																		
-	0	36	19	57	-	23.2	206	96	20	48	40	65	317	150	5.2	1860	220	-	62	42	35	50	399	690	7.3	700	920	-	50	27	19	31	764	993	7.3	400	3520	-
Steady State Period 9 Begins																																					
118	300	60	19	-	-	-	204	110	21	45	40	60	248	79	5.2	1530	310	-	61	42	33	50	482	616	7.1	800	1040	-	51	26	18	29	725	1050	7.5	480	3560	-
119	250	60	19	62	18.6	19.3	205	128	21	45	36	65	288	126	5.5	1040	340	-	64	42	33	46	432	535	7.4	730	1030	-	51	26	18	32	723	1080	7.5	480	3540	-
120	240	60	21	-	-	-	206	127	22	44	35	64	267	135	5.5	1100	210	-	64	40	31	40	489	590	7.4	340	840	-	51	34	22	40	692	914	7.4	350	3570	-
121	300	60	21	-	-	-	205	128	21	38	31	53	210	120	5.5	1250	260	-	64	35	28	42	512	460	7.3	490	880	-	51	27	19	30	685	836	7.4	370	3650	-
122	300	60	18	60	18.6	19.3	208	128	19	44	36	61	252	99	5.4	1190	310	-	63	33	26	40	434	337	7.5	440	960	-	51	26	18	30	719	1040	7.6	290	3580	-
123	280	60	20	60	18.6	19.2	205	128	21	33	26	50	289	80	5.7	800	320	4.8e8	63	31	25	40	352	450	7.4	430	1080	<1e3	51	28	20	31	748	987	7.5	550	3220	<1e3
124	300	60	21	-	-	-	205	134	20	43	35	58	298	120	5.3	1200	210	-	63	29	23	35	478	451	7.3	170	830	-	51	32	20	30	602	826	7.5	610	3080	-
125	300	60	21	61	18.6	19.0	204	127	22	34	27	42	220	88	5.3	1160	390	-	63	32	24	32	456	390	7.2	220	630	-	51	37	25	38	687	860	7.3	920	2810	-
126	330	60	22	61	18.5	19.4	205	126	23	34	27	43	288	108	5.5	810	270	5.2e9	64	25	20	40	403	423	7.1	250	440	8.8e3	51	36	21	34	683	837	7.3	840	2970	<1e3
127	340	60	22	-	-	-	207	128	23	50	41	53	298	122	5.3	1040	210	-	64	28	22	35	398	456	7.1	350	670	-	51	33	23	30	629	824	7.1	920	2210	-
128	340	60	22	-	-	-	206	126	20	39	32	53	210	69	5.0	1600	40	-	64	34	26	36	434	554	7.2	170	630	-	51	32	22	34	628	725	7.0	980	2370	-
129	350	60	21	61	18.6	19.0	205	126	20	42	35	61	246	132	5.2	1170	220	-	64	27	22	32	460	561	7.2	290	790	-	52	31	22	33	625	748	7.2	900	2160	-
Steady State Period 9 Ends																			Average Values for the Last 7 days of Period																		
-	320	60	21	61	18.6	19.2	205	128	21	39	32	51	264	103	5.3	1110	240	-	64	30	23	64	426	469	7.2	270	720	-	51	33	22	33	657	829	7.3	820	2690	-

Table 5a.6 Operational, Chemical and Bacteriological Data for the Dual Digestion Plant Athlone: Phase II Oxygenation with Air + Oxygen

Day	Aeration							Feed Sludge										Aerobic Sludge										Anaerobic Sludge									
No	Q_{AIR}	M_{O2}	T_{in}	T_{out}	O_2^1	O_2^2	I_{mix}	Q_{SL}	T_f	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_r	TS	VS	COD	CON	NH_4	pH	VA	BA	FC	T_d	TS	VS	COD	CON	NH_4	pH	VA	BA	FC
	m^3/h	kg/h	°C	°C	%	%	amp	m^3/d	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl	°C	g/l	g/l	g/l	mS/m	mg/l	-	mg/l	mg/l	/cl
Steady State Period 10 Begins																																					
130	340	96	21	-	-	-	205	158	21	44	36	59	265	85	5.0	1670	90	-	62	30	24	37	198	589	6.8	240	320	-	52	34	25	38	614	745	7.0	1080	2240	-
131	360	96	20	59	18.6	19.3	205	192	21	41	34	59	300	100	4.9	2080	0	1.6e9	63	33	26	38	377	567	7.0	370	430	4.0e3	52	31	21	32	593	724	6.9	1260	1890	<1e3
132	330	96	22	60	18.8	19.2	205	192	24	36	30	56	383	115	4.8	1470	0	-	63	32	26	38	462	541	7.0	320	420	-	52	24	18	29	530	689	7.0	870	2170	-
133	350	96	24	59	18.7	18.8	206	192	23	41	33	58	256	160	5.2	1600	60	-	63	34	29	42	219	432	6.7	570	340	-	53	26	18	29	556	536	6.9	1040	1840	-
134	350	96	23	-	-	-	205	193	24	40	32	58	299	129	5.1	1760	120	-	63	36	28	42	342	344	6.9	430	460	-	53	31	23	34	658	632	6.8	990	2160	-
135	360	96	21	-	-	-	205	192	21	44	37	67	323	143	5.2	1980	190	-	64	35	28	43	331	363	6.9	330	510	-	53	26	21	31	660	545	6.9	890	1970	-
136	370	96	21	60	18.6	19.2	205	192	22	39	33	56	231	161	5.2	1230	280	-	63	35	25	39	372	375	6.8	430	410	-	53	29	20	29	593	663	7.0	680	2340	-
137	360	96	26	61	18.5	19.4	206	191	25	48	39	67	210	120	5.2	1880	210	6.0e8	63	27	23	35	292	462	7.0	240	280	<1e3	53	39	27	40	618	654	6.9	500	2840	<1e3
138	370	96	25	61	18.6	19.0	205	192	24	45	35	64	286	138	5.3	1560	240	2.4e9	64	30	24	38	392	365	6.9	310	560	<1e3	53	30	21	32	589	760	7.0	320	2430	<1e3
Steady State Period 10 Ends																			Average Values for the Last 7 days of Period																			
-	356	96	23	60	18.6	19.1	205	192	23	42	34	61	284	138	5.1	1640	160	-	63	33	26	40	344	412	6.9	380	430	-	53	29	21	32	600	639	6.9	750	2250	-
Steady State Period 11 Begins																																					
139	340	87	24	-	-	-	203	192	26	43	35	64	254	101	5.5	1250	260	1.9e9	62	34	26	42	292	364	6.8	350	690	2.3e4	53	32	22	33	563	689	7.1	450	2880	6.0e3
140	340	76	23	-	-	-	202	191	24	37	29	55	232	126	5.0	1980	80	-	61	36	28	46	210	463	7.0	320	380	-	53	34	24	37	586	758	7.0	340	2770	-
141	360	76	23	-	-	-	203	192	25	46	38	49	254	143	5.2	1880	160	-	61	36	28	40	243	365	6.8	380	430	-	53	36	26	39	605	673	7.2	420	2590	-
142	360	56	24	-	-	-	203	154	25	41	34	51	289	117	5.5	1200	300	-	60	38	30	40	383	325	7.0	320	630		53	29	20	30	585	661	7.1	320	2640	-
143	360	46	23	-	-	-	205	162	25	37	30	56	330	120	5.0	2100	80	1.5e8	58	36	29	44	344	406	6.8	360	500	1.0e5	53	39	26	40	556	752	7.2	480	2730	1.1e4
144	370	76	25	-	-	-	203	192	25	44	34	60	274	78	5.2	1770	330	1.1e9	60	32	25	38	325	385	6.9	290	560	5.0e5	53	33	24	37	591	680	7.2	300	2620	1.4e4
145	380	76	25	56	18.6	18.6	204	192	25	45	34	57	289	86	5.1	1850	40	-	60	37	30	43	209	275	6.9	310	370	-	53	30	22	32	566	721	7.3	480	2840	-
146	400	76	25	-	-	-	203	193	26	36	28	50	271	46	4.9	2220	50	2.0e9	60	38	29	46	355	225	6.7	430	460	2.3e5	53	30	22	32	589	690	7.2	520	2850	1.0e4
147	410	76	25	57	18.6	19.3	205	192	26	42	34	62	265	98	4.8	2310	80	-	60	42	32	49	219	399	6.8	320	500	-	53	28	21	33	558	697	7.4	580	2910	-
148	420	76	23	-	-	-	203	192	23	44	36	69	201	30	4.8	2210	0	-	61	36	28	44	335	357	7.0	350	530	-	53	29	21	30	563	685	7.3	430	2910	-
149	430	76	22	-	-	-	202	191	24	36	30	55	289	120	5.0	2100	210	7.0e8	60	36	28	44	355	312	7.0	230	500	4.8e5	53	29	21	31	525	724	7.2	420	2510	<1e3
150	410	76	26	58	18.5	18.8	203	192	26	42	35	63	208	74	5.5	920	180	1.4e9	60	32	26	35	360	300	6.8	460	550	3.2e5	54	32	23	35	495	664	7.2	460	2830	<1e3
151	420	76	22	58	18.6	18.9	204	191	24	40	33	56	254	90	5.4	1420	270	2.1e9	61	37	30	50	178	347	6.9	170	310	4.2e5	54	25	18	29	516	613	7.4	420	2970	<1e3
152	420	76	23	57	18.6	19.0	205	192	23	47	38	60	234	89	5.3	1320	310	-	60	38	29	41	267	263	6.8	260	490	-	54	26	19	29	523	658	7.3	480	2690	-
Steady State Period 11 Ends																			Average Values for the Last 7 days of Period																			
-	416	76	24	58	18.6	19.0	204	192	25	41	34	59	246	78	5.1	1790	160	-	60	37	29	44	295	314	6.9	310	480	-	53	29	20	31	538	675	7.3	470	2810	-



APPENDIX 6

CALCULATION OF THE AEROBIC REACTOR CARBON BALANCE

6a.1 Introduction

Appendix 6 presents the calculations for the carbon balance across the aerobic reactor for the following three periods of operation at Athlone:

- Phase I (Oxygenation with air alone) Days 1-312 R_h = 4.4days
- Phase II (Oxygenation with air +pure oxygen) Days 6-117 R_h = 1.9 days
- Phase II (Oxygenation with air +pure oxygen) Days 118-152 R_h = 1.1 days

The above three periods of operation were selected to demonstrate the validity of the carbon balance over a range of retention times.

6a.2 The Aerobic Reactor Carbon Balance

Organic carbon enters the aerobic reactor in the volatile solid component of the feed sludge. Inside the aerobic reactor, thermophilic organisms utilise organic matter as substrate for growth and take up oxygen during respiration. Through the oxidative breakdown of organic matter both carbon dioxide and ammonia are released. A large proportion of this carbon dioxide enters the gas phase and passes from the system in the vent gas. However, a fraction (<10%) remains in solution as bicarbonate as a result of the uptake of H^+ ions in the ammonification of NH_3 (see Section 3.1.9). The carbon balance across the aerobic reactor is therefore as follows:

$$M(VS\,as\,C)_{dest} = M(CO_2 as\,C)_{out} + M(NH_4CO_3 as\,C)_{gen} \quad \text{...kg(C)/d} \quad (6a.1)$$

where:

$M(VSasC)_{dest}$ = mass rate of VS destruction in terms of carbon release (kg(C)/d)

$M(CO_2asC)_{out}$ = mass rate of CO_2 in the reactor vent gas stream (kg(C)/d)

$M(NH_4CO_3asC)_{gen}$ = mass rate of NH_4CO_3 generation in the aerobic reactor (kg(C)/d)

In proportion to the mass of carbon dioxide generated in the system, the mass of carbon dioxide entering the system via the influent air stream considered negligible.

6a.3 The Rate of Volatile Solids Destruction

The average rate of volatile solids destruction $M(VS)_{dest}$ in units of kg(VS)/h was measured during both phase I and phase II of the investigation (refer to Sections 4.4.1 and 7.4.2 respectively). The results obtained are presented in Table 6a.1 below.

Accepting a $C_5H_7O_2N$ formula to represent the volatile soiid content of the feed sludge, with a stoichiometric content of 53%, the daily rate of carbon release $M(VSasC)_{dest}$ through the oxidation of organic matter is therefore given by:

$$M(VS\,as\,C)_{dest} = 24 \times 0.53 \times M(VS)_{dest} \quad \text{...kg(C)/d} \quad (6a.2)$$

The rate of carbon release for the three periods is therefore as follows:

Table 6a.1 The Average Rates of Volatile Solids Destruction Across the Aerobic Reactor During Phase I and Phase II in units of kg(VS)/h and Kg(C)/d

Phase	Days	Oxygenation Mode	Retention Time (d) R_h	Rate of VS Dest. kg(VS)/h $M(VS)_{dest}$	Rate of Carbon Release kg(C)/d $M(VSasC)_{dest}$
I	1-312	Air	4.4	16.2	**206**
II	6-117	Air+Oxygen†	1.9	25.1	**319**
II	118-152	Air+Oxygen	1.1	47.0	**598**

† Excludes days 41-73 when recirculation line was o/c

6a.4 The Rate of Carbon Dioxide Generation

The average rate of oxygen utilisation $M(O_2)_{ut}$ in units of kg(O_2)/h was measured during both phase I and phase II of the investigation (refer Sections 4.4.2 and 7.4.2 respectively). The results obtained are presented in Table 6a.2 below.

The respiration quotient Y_{CO2}, defined as the number of moles of CO_2 generated per mole of O_2 utilised was measured at 0.70 mol/mol during phase I (refer Section 4.3.6). This figure was accepted to be suitable for use during phase II (as an appropriate instrument was no longer available to measure the vent gas CO_2 concentration). The correctness of the measured Y_{CO2} value was confirmed by, (1) the good agreement with the Y_{CO2} value (0.67 mol/mol) measured by Messenger *et al* (1992) and, (2) the accuracy of the steady state heat balances compiled during phase II, which yielded an identical value for the specific heat yield coefficient of Y_h=12.8MJ/kg(O_2).

In terms of mass, the respiration quotient corresponds to (0.70 x 44/32) 0.963 kg(CO_2)/kg(O_2). With the carbon content of carbon dioxide being 27%, the daily rate of carbon loss from the aerobic reactor in the vent gas is given by:

$$M(CO_2 as C)_{dest} = 24 \times 0.27 \times 0.963 \times M(O_2)_{ut} \quad \text{...kg(C)/d} \quad (6a.3)$$

The rate of carbon loss in the vent gas for the three periods is therefore as follows:

Table 6a.2 The Average Rates of Carbon Loss in the Vent Gas (due to CO_2 Emission) from the Aerobic Reactor During Phase I and Phase II in units of kg(C)/d

Phase	Days	Oxygenation Mode	Retention Time (d) R_h	Rate of Oxygen Util. kg(O_2)/h $M(O_2)_{ut}$	Rate of C Loss in Vent Gas kg(C)/d $M(CO_2 asC)_{gent}$
I	1-312	Air	4.4	27.6	**174**
II	6-117	Air+Oxygen†	1.9	34.3	**216**
II	118-152	Air+Oxygen	1.1	75.7	**478**

† Excludes days 41-73 when recirculation line and pump were not operational

6a.5 The Rate of Ammonium Bicarbonate Generation

The average increase in ammonium ion concentration after aerobic treatment (symbolised here as ΔNH_4 in units of g(N)/m^3) was measured during both phase I and phase II of the investigation (refer to Sections 4.4.3 and 7.4.2 respectively). The results obtained are presented in Table 6a.3 below.

The reaction between ammonia and carbon dioxide is as follows:

$$NH_3 + CO_2 + H_2O \rightarrow NH_4HCO_3$$

In terms of the above reaction equation, 14g of nitrogen (17g of ammonia) will react with 12g of carbon (44g of carbon dioxide) to form 78g of ammonium bicarbonate. With an influent sludge flow rate $Q(SL)_{in}$ (in units of m^3/d), the rate of uptake of carbon to form ammonium bicarbonate $M(NH_4CO_3 asC)_{gen}$ is therefore given by:

$$M(NH_4HCO_3 as C)_{gen} = Q(SL)_{in} \times \frac{12}{14} \times \frac{\Delta NH_4}{1000} \quad \text{...kg(C)/d} \quad (6a.4)$$

The rate of uptake of carbon into the liquid phase for the three periods is therefore as follows:

Table 6a.3 **The Average Rates of Carbon Loss in the Liquid Phase (due to NH_4CO_3 Formation) from the Aerobic Reactor During Phase I and Phase II in units of kg(C)/d**

Phase	Days	Oxygenation Mode	Retention Time R_h -d (Sludge Flow $Q(SL)_{in}$ m^3/d)	Increase in NH_4 Concentration $g(N)/m^3$ ΔNH_4	Rate of C Loss in Liquid Phase kg(C)/d $M(CO_2asC)_{gent}$
I	1-312	Air	4.4 (41)	252	**9**
II	6-117	Air+Oxygen†	1.9 (95)	480	**39**
II	118-152	Air+Oxygen	1.1 (152)	311	**45**

† Excludes days 41-73 when recirculation line was o/c

Messenger *et al* (1992) applied fundamental water chemistry principles to calculate the dissolved CO_2 concentration in the reactor effluent sludge stream from the reactor partial pressure of CO_2 and a calculated insitu pH from measured $H_2CO_3^*$ alkalinities. They did this to check whether or not the dissolution of CO_2 in the liquid phase significantly affects the respiration quotient. They found that it would at most increase by 5% i.e. from 0.66 to 0.70 mol(CO_2) generated per mol(O_2) utilised. Because the proportion of CO_2 leaving the reactor is relatively so small, it was considered sufficiently accurate to estimate it from simple stoichiometry (Table 6a.3) rather than the complex water chemistry approach.

6a.6 Calculation of the Carbon Balance

The carbon balance as defined by Eq 6a.1 above can now be examined. From the data tabulated in Tables 6a.1 to 6a.3 the carbon balances for the three periods are presented in Table 6a.4 below:

Table 6a.4 Results of the Carbon Balance for Three Distinct Operating Periods For the Aerobic Reactor - Athlone: Carbon Flows as kg(C)/d

No	Carbon Release	Carbon Loss		Error (%)
	$M(VSasC)_{dest}$	$M(CO_2asC)_{gen}$	$M(NH_4CO_3asC)_{gen}$	
1	206	174	9	89
2	319	216	39	80
3	598	478	45	87

Note: 1 Phase I (days 1-312) Oxygenation with air alone R_h=4.4days

2 Phase II (days 6-117) Oxygenation with air + oxygen R_h=1.9days

3 Phase II (days 118-152) Oxygenation with air + oxygen R_h=1.1days

6a.7 Discussion

For operation of the Athlone aerobic reactor, whether with air or a combination of air plus pure oxygen, at retention times ranging from 1 to 6 days the carbon balance shows a +80% agreement (see Table 6a.4 above) i.e. of the carbon VS (assumed to be $C_5H_7O_2N$) entering the system via the sludge liquid stream, 80-89% could be measured leaving the system in the gas and liquid phases (as CO_2 and NH_4HCO_3 respectively). Considering the approximation made in calculating the carbon balance (the acceptance of a carbon content of 53% for the feed VS), the balance is considered to be satisfactory; If a carbon content of 47% is taken for the feed VS, the accountability of the carbon approximates to 100%.

During phase II (oxygenation with air plus pure oxygen), with the aerobic reactor operated at relatively short retention times (1 to 2 days), the rate of volatile solids destruction was found to be linked to the rate of oxygen utilisation rate (see Section 7.5). Whilst insufficient data was available to make a direct claim to linearity, the success of the carbon balance provides support for the claim that the two parameters are somehow linked, even at short retention times.

It is interesting to note that Messenger *et al* (1992), on the Milnerton short retention time (1 to 2 days) pure oxygen aerobic system measured minimal VS destruction across the aerobic reactor (1.3%) and no link was observed with the rate of oxygen utilisation. The carbon balance across the Milnerton reactor was 1330% i.e. the mass of carbon dioxide in the vent gas (as C) was 13.3 times greater than that given by VS removal.

REFERENCES

Messenger JR, Ekama GA, de Villiers HA, Kenmuir K and Laubscher SA (1992) Evaluation and optimisation of dual digestion of sewage sludge - Part 2: Aerobic reactor performance. Final report WRC 189/3/92, Water Research Commission, PO Box 824, Pretoria, 0001.

APPENDIX 7

FINAL SLUDGE STABILITY: THE SPECIFIC OUR TEST

7a.1 Introduction

This appendix describes the results of an investigation to test the stability of the final sludge from the dual digestion process by measuring the specific oxygen utilisation rate (*SOUR*) which the sludge stimulates under laboratory batch test conditions. To provide comparison, a number of different sludge types produced at plants in the Western Cape (listed in Table 7a.1 below) were tested in conjunction (More, 1995).

7a.2 Objective of the *SOUR* Test

Wastewater sludges which have undergone treatment for stabilisation (whether it be by anaerobic or aerobic digestion, Zimpro, composting etc...) may still contain residual energy in the form of biodegradable organic matter, which if not further degraded may stimulate biological activity following land disposal.

The fraction (expressed as a percentage) of the feed sludge volatile solids destroyed in anaerobic/aerobic digestion treatment processes has traditionally been used as a performance parameter and as a general indicator of the degree of digestion (stability). Heidman (1989) considers that >38% volatile solids removal is necessary for satisfactory sludge stabilisation. Whilst VS reductions in excess of 38% are readily achievable when treating primary sludges, lower values than 38% can occur if the sludge being treated is principally secondary sludge (or has been partially stabilised by other pre-treatment methods). The final sludge may well be 'fully stabilised' and not comply with the VS reduction criterion; a fact common in the aerobic digestion of secondary sludges. Recently, the parameter has been introduced more formally as a criterion of adequate digestion.

To overcome the problem of sludge stability assessment, the stability of aerobically digested sludges has been tested in terms of the oxygen demand exerted. Bruce and Fisher (1984) consider a specific oxygen utilisation rate (*SOUR*) of < $2mg(O_2)/g(VS).h$ at 18°C to be indicative of a stable sludge. Heidman (1989) quotes the current U.S. Federal Regulations by stating that a *SOUR* of < $1mg(O_2)/g(TSS).h$ should be achieved.

Whilst the ***SOUR*** parameter has been used principally in evaluating the stability of aerobically digested sludges, tests were carried out on the final sludge from the dual digestion process (and other types for comparison) in order to provide further information with regard to final sludge stability.

The problem exists that as the measurement of ***SOUR*** is an aerobic test, sludge samples which have been subjected to anaerobic conditions (as in dual digestion) will require a period of time under aerobic conditions to enable an aerobic population to develop. Consequently, the ***SOUR*** tests in this investigation were performed over a period of 4 days.

7a.3 *SOUR* Experimental Procedure

The batch tests were carried out in a cylindrical vessel with a process capacity of 3 litres (see Figure 7a.1). The contents of the vessel were continuously stirred with a slowly rotating paddle fixed to a centre column. The temperature of the sludge in the vessel was maintained at 20°C. Air was passed into the vessel via a fine bubble diffuser situated at the base of the vessel below the paddle. The D.O. probe was connected to an automated unit (developed by Randall *et al* ,1991) capable of controlling the D.O. concentration between a high and a low set-point and calculating a continual output of ***OUR*** data obtained during the air-off period. The D.O. level in the sludge was maintained between 2.0 and 4.5 mg(O_2)/l by automatically switching on and off the air supply. The operating range of D.O. is sufficiently wide to enable the ***OUR*** to be measured with a high degree of confidence and sufficiently short to enable four ***OUR*** values to be produced per hour. Plastic balls were spread onto the surface of the liquid to minimise foaming and to prevent O_2 absorption from the air. Before testing, sludge samples were sieved to remove the coarse particles. If the sludge was still relatively concentrated (>15kg(TS)/m^3) the sample was diluted with water. The total and volatile suspended solids concentrations and 0.45μm filtered COD and TKN

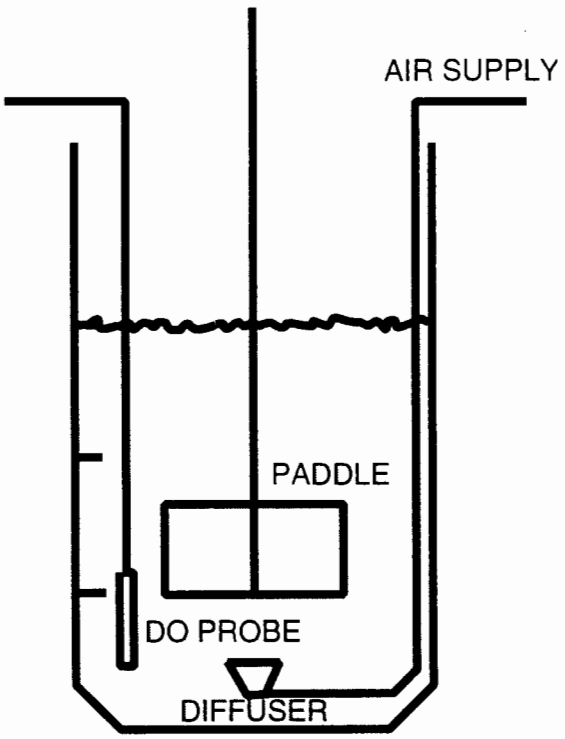


Figure 7a.1 Schematic of the Batch Reactor Employed in the Determination of the *SOUR*

concentrations were determined at regular intervals during the batch test (More, 1995). A graphical display of the *SOUR* data is presented in this Appendix.

7a.4 Presentation of the *SOUR* Data

The various sludge types, and their sample dates, which were subjected to testing for *SOUR* analysis are listed in Table 7a.1 below. The concentrations of TS and VS (g/l) of the samples prior to testing (after sieving and dilution, if necessary) are also given. The variation in *SOUR* with time for each sample is presented in Figure 7a.2 below. A discussion of the results is presented in Section 7a.5 below.

Table 7a.1 Listing of the Various Sludge Types Subjected for *SOUR* Testing, Sample Dates, and Initial TS and VS (g/l) Concentrations

Sludge Type	Sample Date	TS (g/l)	VS (g/l)
Dual Digestion			
- Dual Digester Athlone (SRT=10days)	23/2/95	6.41	4.10
- After Secondary Digestion (SRT=20days)	23/2/95	8.18	5.62
Conventional Anaerobic Digestion			
- Primary Digester Athlone (SRT=20days)	1/3/95	7.30	5.05
Conventional Aerobic Treatment Borchards Quarry			
- Waste Activated Sludge (SRT=20days)	13/3/95	13.48	11.36
- Aerobic Digestion (SRT=20days)	7/3/95	10.79	9.24
Zimpro			
- Milnerton (±200°C)	21/3/95	7.09	5.87

7a.5 Discussion of Results

Dual Digested Sludge

Analysis of the *SOUR* plot (Figure 7a.2a) for the dual digested sludge, shows that in the initial period (0-12hrs) the *SOUR* exhibits an exponential increase, in all likelihood due to the growth of aerobic bacteria utilising soluble readily biodegradable COD (RBCOD). This growth activity continues until ±10hrs (with the *SOUR* peaking at 13gO/kgTSS.h) after which the *SOUR* drops rapidly as a consequence of the exhaustion of the soluble RBCOD. Thereafter a situation develops whereby the bacteria develop and die as a

consequence of (i) the complete utilisation of hydrolysed particulate substrate and/or (ii) predation by other bacteria (an increase in predation will result in an increase in ***SOUR***). From the TKN results, no nitrification took place in these tests. Towards the end of the test, the target ***SOUR*** of 1gO/kgTSS.h is attained.

The ***SOUR*** plot (Figure 7a.2b) for the dual digested sludge after secondary anaerobic digestion, exhibits a similar trend to that described above except that the peak ***SOUR*** (3.5gO/kgTSS.h) occurs at the start of the test and is only a quarter of that observed for the dual digested sludge. From the start of the test the ***SOUR*** drops and is followed by occasional local peaks. While the frequency of these peaks is greater than that observed in Figure 7a.2a, the target ***SOUR*** is achieved at times of ±18, ±30, and ±90hrs.

Activated Sludge

Analysis of the ***SOUR*** plot (Figure 7a.2c) for the waste activated sludge (ex Borchards Quarry Sewage Works - SRT 20days) shows a sharp increase in the ***SOUR*** in the initial stages of the test (<2hrs, not visible in Figure 7a.2c) due to the growth of aerobic bacteria utilising RBCOD (peak value 13gO/kgTSS.h). Following which, in a time space of ±3hrs, there is an exponential decrease due to the depletion of the available substrate. A target ***SOUR*** of 1gO/kgTSS.h is achieved after ±20hrs. Thereafter a gradual decrease in ***SOUR*** is observed. It is noted that no local peak points occur, indicating that the waste activated sludge is relatively well stabilised containing little trapped energy in the form of biodegradable organic matter.

Analysis of the ***SOUR*** plot (Figure 7a.2d) for the same waste activated sludge after aerobic digestion (SRT 20days) shows a similar plot to that described above. The initial peak is significantly lower (3gO/kgTSS.h) and the target ***SOUR*** is attained as early as ±2hrs. Again a gradual decrease in ***SOUR*** is observed with levels being lower than that recorded for the waste activated sludge. The general appearance of the plot is consistent with the fact that the activated sludge has undergone further stabilisation in the aerobic digester and therefore contains less biologically utilisable energy.

Conventional Anaerobic Digestion

The ***SOUR*** plot (Figure 7a.2e) for the anaerobically digested sludge is similar in appearance to that observed for the dual digested sludge (Figure 7a.a). Both plots show an initial peak ***SOUR*** of similar height (12-13 gO/kgTSS.h). However, the peak for the anaerobic sludge does take longer to generate (±15hrs as apposed to 10hrs). This may be an indication that the population of facultative aerobic/anaerobic bacteria is lower than in the dual digested sludge. This could be a consequence of the dual digested sludge having undergone thermophilic aerobic pre-treatment.

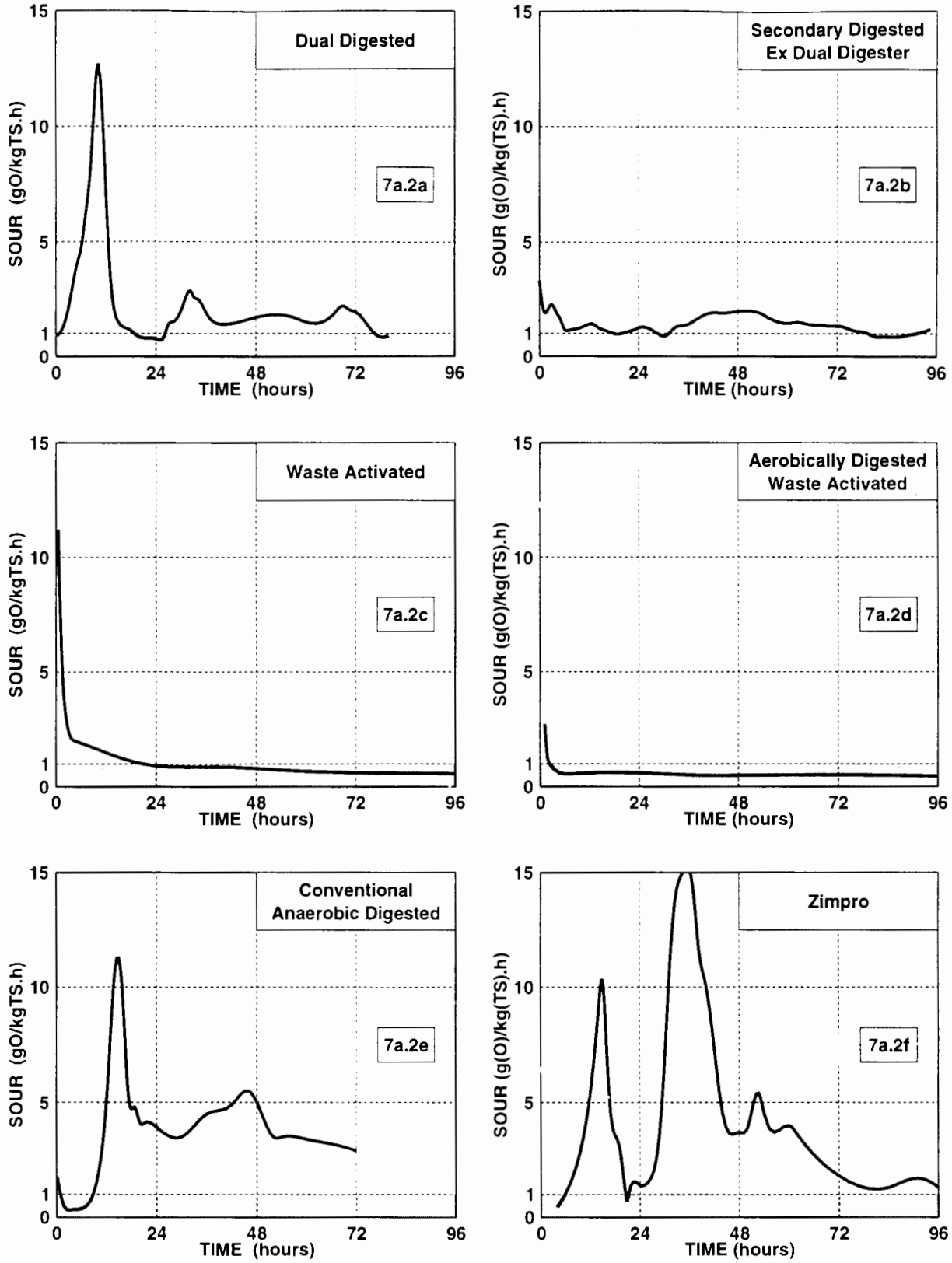


Figure 7a.2 Sludge Stability Tests: The Variation in *SOUR* (gO/kgTSS.h) with Time (0-96hrs) for Different Effluent Sludge Types

After the initial peak, a number of local peaks are observed. These occur at different times to that observed for the dual digested sludge, again a possible indicator of a

difference in bacterial population. The general background levels and final level (±3gO/kgTSS.h) for the ***SOUR*** are higher than that encountered for the dual digested sludge, indicating a lesser degree of stability. At no point during the test, after the growth of an adequate population of aerobic bacteria, did the ***SOUR*** reach the target value of 1gO/kgTSS.h.

Zimpro Sludge

The ***SOUR*** plot (Figure 7a.2f) for the Zimpro sludge is in complete contrast to the others. At the beginning of the test exponential growth is observed, reaching a peak at ±10gO/kgTSS.h. This however is followed by an even larger peak which occurs at ±36hrs (peak value ±15gO/kgTSS.h). Thereafter the ***SOUR*** falls off, as available substrate is utilised.

The results are unexpected in that during the Zimpro process the sludge is subjected to very high temperatures (±200°C). Consequently, it was considered that this would be the most stable of all the sludges tested. However it produced the worst results. Except for the start of the batch test, and briefly between major peaks, at no other time did the ***SOUR*** reach the target value of 1gO/kgTSS.h.

Erratic ***SOUR*** peaks were observed in some of the earlier batch tests and were found to be due to foaming; foam collected on the surface plastic balls and reactor walls. The foam was brushed back into the reactor once daily at about 24h intervals. In some of the tests (not shown in this appendix), the ***SOUR*** increased precipitously upon brushing the reactor, surface balls and walls. The increases in ***SOUR*** for the Zimpro sludge are not precipitous, nor at approximate 24 hour intervals, and therefore it was concluded that the ***SOUR*** peaks for this sludge are due to biological activity within the sludge.

7a.6 Conclusions

The quantity of oxygen utilised during the first 24 hours, for each of the sludge types tested, is tabulated in Table 7a.2 below. The average ***SOUR***'s for this initial 24hr period are quoted. The only sludge which meets the 1gO/kgTSS.h target is the aerobically digested waste activated sludge from Borchard's Quarry.

The ***SOUR*** test is ideally suited to the testing of aerobic sludges. In the case of the anaerobic and Zimpro sludges spurious peaks are observed during the batch test due to the growth/die-off/predation of aerobic bacteria and reactor cleaning making it difficult to quantify a result in terms of ***SOUR***; No specific periods could be defined at which it could be asserted that the sludge is stable in terms of the target ***SOUR*** of 1gO/kgTSS.h.

Table 7a.2 *SOUR* Sludge Stability Test: Oxygen Utilisation (gO/kgTSS) During the First 24hrs of the Batch Test and the Average *SOUR*'s Recorded During this Period (gO/kgTSS.h) for the Different Sludge Types Tested

Sludge Type	Oxygen Utilised During First 24hrs gO/kgTSS	Average *SOUR* During First 24hr Period gO/kgTSS.h
Dual Digested Sludge	81.6	3.4
- After Secondary Digestion	33.1	1.4
Waste Activated Sludge	46.6	1.9
- After Aerobic Digestion	15.6	0.7
Conventional Anaerobic Digested Sludge	86.4	3.6
Zimpro Sludge	68.6	2.9

The stability of the sludge from the Dual Digester, in terms of the *SOUR* test, does however compare favourably with that observed for conventionally digested sludge. This inspite of the fact that the solids retention time through the dual digestion process is half that through the conventional process; a likely consequence of both stages in the dual digestion process operating at thermophilic temperatures (biological reactions proceed much faster at higher temperatures) and the conditioning effect of aerobic pre-treatment. This provides important information on the stability of short retention dual digester sludge. However, not only does this aspect require further investigation, but also the batch test itself requires further study.

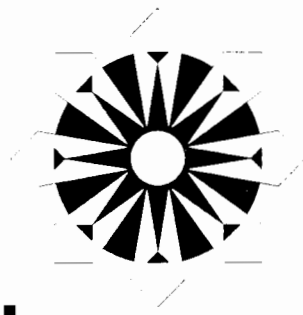
REFERENCES

Bruce AM and Fisher WJ (1985) Sludge stabilisation - methods and measurement. In: *Sewage Sludge Disinfection and Stabilisation*, Edited by Bruce AM, Ellis Harwood Publishers Ltd., Chichester, England.

Heidman JA (1989) Aerobic thermophilic sludge digestion in the Federal Republic of Germany - Chapter 4 Regulatory Background, EPA Report - Contract No. 68-03-3429

More LM (1995) Sludge stability assessment by means of aerobic batch tests. BSc thesis. Dept. of Civil Eng., Univ. of Cape Town, Rondebosch, South Africa.

Randall EW, Wilkinson A, and Ekama GA (1991) An Instrument for the direct determination of oxygen utilisation rate. *Water SA* **17**(1) 11-18



APPENDIX 8

COMPUTER PROGRAMMES FOR CALCULATING THE STEADY STATE HEAT BALANCE

This appendix gives a brief description and listing of three of the principle computer programmes compiled during the course of this investigation. Each of the programmes were written in the C programming language for use on the UNIX-DOS computer operating system. The three programmes are:

1 Calculation of the Oxygenation and Heating Terms - Phase I
2 Prediction of Aerobic Reactor Operating Temperature - Phase I
3 The General Dual Digestion Simulation Model (described in Chapter 8)

8a.1 Programme 1: Calculation of the Oxygenation and Heating Terms - Phase I

Programme 1 calculates the oxygenation and heating terms required to solve the steady state heat balance for phase I (oxygenation of the aerobic reactor with air alone). Details regarding compiling the programme, the required format of the input data file and the command line instruction are given in the programme. Giving the appropriate operating data, the programme will compute values for; (1) the oxygen supply rate, transfer rate and transfer efficiency; (2) the respiration quotient; (3) the terms in the steady state heat balance; and (4) the specific heat yield coefficient. The equations used in the programme were derived in Chapter 3. A similar programme was compiled to calculate the oxygenation and heating terms during phase II (oxygenation with both air and pure oxygen).

Listing of Programme 1

```
/* DUAL DIGESTION */
/* COMPUTER PROGRAMME FOR THE CALCULATION */
/* OF (1) THE AERATION CHARACTERISTICS    */
/*    (2) THE RESPIRATION QUOTIENT        */
/*    (3) THE TERMS IN THE HEAT BALANCE   */
/*    (4) THE SPECIFIC HEAT YIELD         */
/* Compile Command "cc -o progname -lm cprog.c"*/
/* Command Line "progname inputdatafile > outputdatafile"*/
/* Required Format of Input Data File */
/* Date  QSLi  TSLi  TSLo  TSLd  TAIRi  QAIRi  O2  CO2  TAIRo */
/* Format of Output Data File */
/* Date  Hm Hb Hs Hg Hv Hw Hi Ho Yh OSR OTR OTE O2 CO2 N2 YCO2 */
#include <stdio.h>
#include <ctype.h>
#include <math.h>
#define  YES    1
#define  NO     0
#define  MVAL   -1  /* missing numeric value */
#define  MAXFLDS 48
double sqrt();
double pow();
int    getData();
char   pgmname[] = {"heats"};
char   oprec[1024];
char   iprec[1024];
int    ln;
main(argc,argv)
  int    argc;
  char  *argv[];
  {
    FILE   *openFile();
    FILE   *fp;             /* input file ptr */
    char   *sp;
    while ( --argc && (*++argv)[0] == '-' )
      {  /* Interpret cmd line */
        sp = argv[0] + 1;
        switch(*sp)
          {
            case '-':
              dspUsage();  exit(1);
              break;
            default:
              fprintf(stderr,"%s: illegal option -%c\n",pgmname,*sp);
              dspUsage();
              exit(1);
              break;
          }
      }
    if ( argc > 1 )
```

```
    { /* extra arguments of wrong kind */
      fprintf(stderr,"Too many arguments\n");
      dspUsage();
      exit(1);
    }
  else if ( (fp = openFile(argc,argv)) == NULL )
    exit(1);  /* can't open ip file */

      /* The next fn does the hard work */
  calcHeats(fp);
  exit(0);
 }

/***/ calcHeats(fp)
 FILE *fp;
 {
 int w,d;  /* format control variables */
/* HEAT BALANCE - DUAL DIGESTER */
double OSR,OTR,OTE,QAIRi,O2,Pvapo,Pvapi,TAIRo,TAIRi,Uo,Ui,Hg,Hs,TSLo,TSLi,QSLi;
double Hv,Hw,TSLd,Hm,Hb,Hi,Ho,N2,CO2,Yhcalc,Yh,YCO2;
printf("Date     H(m) H(b) H(s) H(g) H(v) H(w) H(i) H(o) Yh OSR OTR OTE O2 CO2 YCO2\n");
   setDestPtr(oprec);
   while (getData(fp,&QSLi,&TSLi,&TSLo,&TSLd,&TAIRi,&QAIRi,&O2,&CO2,&TAIRo))
     {
/* CALCULATION OF THE AERATION CHARACTERISTICS */
Yh    = 13.20;
OSR   = (0.00152*QAIRi);
OTR   = (QAIRi*(21.00-O2))/(138.00*(100.00-O2+(0.70*O2)));
OTE   = (OTR/OSR)*(100.000);
/* CALCULATION OF THE VAPOUR HEAT LOSS IN THE EFFLUENT GAS */
Pvapo = pow(10,(8.903-(2239.0/(273.0+TAIRo))));
Pvapi = pow(10,(9.124-(2307.0/(273.0+TAIRi))));
Uo    = 18.00*Pvapo/(29.00*(760-Pvapo));
Ui    = 18.00*Pvapi/(29.00*(1240-Pvapi));
Hv    = 2.382*1.205*QAIRi*(Uo-Ui);
/* CALCULATION OF SENSIBLE HEAT LOSS IN THE EFFLUENT GAS */
Hg    = ((1.205*QAIRi)/1000.00)*((1.00*(TAIRo-TAIRi))+(Uo*1.870*TAIRo));
/* CALCULATION OF SENSIBLE HEAT LOSS IN THE SLUDGE */
Hs    = (4/24)*((TSLo * QSLi)-((1.205*TSLo*QAIRi*Uo*24)/1000.00)-(QSLi*TSLi));
/* CALCULATION OF WALL HEAT LOSS */
Hw    = (0.820*(TSLo-TSLd))+(0.490*(TSLo-TAIRi));
/* CALCULATION OF MECHANICAL HEAT INPUT */
Hm    = 1.65*19.00;
/* CALCULATION OF BIOLOGICAL HEAT */
Hb = Yh*OTR*184.00;
/* OVERALL HEAT FLOWS */
Hi = Hb + Hm;
Ho = Hv + Hs + Hw + Hg;
/* EFFLUENT GAS CONCENTRATIONS */
N2 =  100-(O2+CO2);
/* RESPIRATION COEFFICIENT */
```

```
YCO2 = CO2/(26.58-(1.2658*O2)-(0.2658*CO2));
Yhcalc = (Ho-Hm) / (OTR*184.00);

        fmtFld("%4.0f ",Hm);
        fmtFld("%4.0f ",Hb);
        fmtFld("%4.0f ",Hs);
        fmtFld("%4.0f ",Hg);
        fmtFld("%4.0f ",Hv);
        fmtFld("%4.0f ",Hw);
        fmtFld("%4.0f ",Hi);
        fmtFld("%4.0f ",Ho);
        fmtFld("%4.1f ",Yhcalc);
        fmtFld("%4.3f ",OSR);
        fmtFld("%4.3f ",OTR);
        fmtFld("%4.3f ",OTE);
        fmtFld("%4.1f ",O2);
        fmtFld("%4.1f ",CO2);
        fmtFld("%4.1f ",N2);
        fmtFld("%4.3f ",YCO2);

        putNewLine();
        printRec(oprec,ln);
        setDestPtr(oprec);
     }
   return(1);
 }

/***/ int  getData(fp,QSLip,TSLip,TSLop,TSLdp,TAIRip,QAIRip,O2p,CO2p,TAIRop)
 /* Reads in data from input file fp & */
  FILE   *fp;
  double *QSLip,*TSLip,*TSLop,*TSLdp,*TAIRip,*QAIRip,*O2p,*CO2p,*TAIRop;
  {
    double v;
    double *dp;
    int    jj, yn;
    ln = 0;
    if ( ! getRec(fp,iprec) )
      return(0);
    ln++;
    setSourcePtr(iprec);
    fndNextFld();
    mvFld();
    if (getNumFld("%lf", QSLip)  != 1 ||
        getNumFld("%lf", TSLip)  != 1 ||
        getNumFld("%lf", TSLop)  != 1 ||
        getNumFld("%lf", TSLdp)  != 1 ||
        getNumFld("%lf", TAIRip) != 1 ||
        getNumFld("%lf", QAIRip) != 1 ||
        getNumFld("%lf", O2p)    != 1 ||
        getNumFld("%lf", CO2p)   != 1 ||
        getNumFld("%lf", TAIRop) != 1
```

```
    )
    { fprintf(stderr,"%s: invalid data: %s\n",pgmname,iprec);
      exit(1);
    }
  yn = fndNextFld();
  if (yn == 'y')
    {
    }
  else
    {
    }
  return(1);
}

/***/  dspUsage()
 {
  fprintf(stderr,"Usage: %s  [-axx] [-byy]  inputfile\n\n",pgmname);
  fprintf(stderr,"          --    displays usage \n");
 }

/***/ FILE *openFile(argc,argv)
 /* Opens file, with name given in *argv, for reading.
    Returns:  Ptr to file if name given else  ptr to stdin.
              Null if named file can't be opened.
 */
 int   argc;
 char *argv[];
 {
  FILE *fp;
  if ( argc == 0 )
    return(stdin);
   if ( ( fp = fopen(argv[0],"r") ) == NULL )
      fprintf(stderr,"%s: can't open %s\n",pgmname,argv[0]);
  return(fp);
 }

/***/ getRec(fp,rec)
 /* Inputs a record ended by \n from file fp. Replaces the \n by \0.
    Returns: Yes if done, No if eof.
 */
 FILE  *fp;
 char  *rec;
 {
  char *sp;
  int   c;
  for ( sp = rec; (c = getc(fp)) != '\n' && c != EOF;  )
    *sp++ = c;
  *sp = '\0';
  return( c == EOF ? NO : YES );
 }
```

```
/***/ printRec(oprec,ln)
 /* Sends oprec to stdout.
 */
  char *oprec;
  int   ln;
  { int nb;
    nb = strlen(oprec);
    if ( fwrite(oprec,sizeof(*oprec),nb,stdout) != nb )
      fprintf(stderr,"correct: write error on line %d.\n",ln);
  }

/***/ movmem(from,to,nch)
  char *from, *to;
  int   nch;
  {
    while (nch--)
      *to++ = *from++;
  }

/*** Functions for Manipulating Fields in a Rec (string) ***/

 /* Fields consist of non-blank chrs and are separated by blanks.
    Blanks are spaces or tabs.
    Descriptive flds start with an alphabetic char.
    Numeric fields contain no alpha chars.
    Missing numeric flds contain m's.
    Two global ptrs are used:
       sp is ptr to source string,
       dp is ptr to destination string.
    The various fns move these along the strings and leave them in
    the correct places after being invoked.
    Ptr sp and dp can be initialised by setSourcPtr() & setDestPtr().
 */
char  *sp;  /* source pointer */
char  *dp;  /* destination pointer */

/***/ setSourcePtr(rec)
  /* Initialises global ptr, sp.
  */
  char  *rec;
  { sp = rec; }

/***/ setDestPtr(rec)
 /* Initialises global ptr, dp.
 */
  char  *rec;
  { dp = rec; }

/***/ chSourcePtr(ch)
 /* Changes global ptr, sp, by the amount in ch.
 */
```

```
 int ch;
 {
   sp += ch;
 }

/***/ fndNextFld()
 /* Moves ptr, sp, to start of next fld in string.
    Returns: first char of fld, or 0, if end of string.
 */
 {
   while ( *sp == ' ' || *sp == '\t' )
     sp++;
   return( *sp );
 }

/***/ mvFld()
 /* Moves fld pointed to by sp to posn pointed to by dp.
    Also moves blanks after the fld.
    sp & dp are left pointing to the start of the next fld.
    leaves 0 at dp posn.
 */
 {
   while ( *sp != ' ' && *sp != '\t' && *sp != 0 )
     *dp++ = *sp++;
   while ( *sp == ' ' || *sp == '\t' )
     *dp++ = *sp++;
   *dp = '\0';
 }

/***/ getNumFld(fmt,np)
 /* Searches string ( global ptr sp ) for the next numeric field,
    either filled or missing ( ie starting with 'm' ),
    converts it according to the control string, fmt (see sscanf()),
    and stores result in *np.
    Can be used for any numeric type by supplying the appropriate
    fmt & np. eg "%ld" & double *np.
    Leaves sp pointing to the first char after the numeric field.
    Returns: YES if done, NO if # not found before end of str,
             MVAL if missing.
 */
 char *fmt;
 {
   int result;
   for ( result = 0; result != 1 && *sp != 0;  )
     {
       while ( *sp == ' ' || *sp == '\t' )
         sp++;
       if ( *sp )
         {
           if (*sp == 'm')
             {
```

```
            mvpastFld();
            return(MVAL);
          }
        result = sscanf(sp,fmt,np);
        mvpastFld();
      }
    }
  return(result);
 }

/***/ fmtFld(fmtcntrlp,val1,val2,val3,val4,val5)
 /* Converts vals to string form using the string fmtcntrlp for
    format control (see printf()), and placing the result at the
    posn given by dp.  Leaves dp just beyond the result, with 0
    at that posn.
    The types and numbers of vals can vary but must agree with
    the format control string.
 */
 char *fmtcntrlp;
 {
  sprintf(dp,fmtcntrlp,val1,val2,val3,val4,val5);
  while (*dp)
    dp++;
 }

/***/ putNewLine()
 {
  *dp++ = '\n';
  *dp   = '\0';
 }

/***/ fndnthFld(n)
 /* Moves global ptr sp to start of nth field (count starts with 1).
    Returns: YES if found, NO if not found before end of string.
 */
 int n;
 {
  while(1)
    {
      while ( *sp == ' ' || *sp == '\t' )
        sp++;  /* find start of string */
      if ( *sp == 0 || --n == 0 )
        return(n == 0);
      else
        mvpastFld();
    }
 }

/***/ fndDescFld()
 /* Moves ptr sp to start of first available descriptive fld ie fld
    starting with alpha char (not 'm').
```

```
   Returns: First chr of fld if found, 0 if none before end of string,
   or before first numeric value or missing value found.
*/
 {
   while(1)
     {
       while ( *sp == ' ' || *sp == '\t' )
         sp++;
       if ( *sp == 0 || *sp == 'm' )
         return(NO);
       else if ( isalpha(*sp) )
         return(*sp);
       else
         return(NO);
     }
 }

/***/ fndNumFld()
 /* Moves sp to start of next available numeric fld in string.
   Returns: YES if found, NO if not found before end of string,
            MVAL if fld starting with 'm' found.

*/
 {
   while (1)
   {
     while ( *sp == ' ' || *sp == '\t' )
       sp++;
     if ( isdigit(*sp) || *sp == '-' || *sp == '+' )
       return(YES);
     else if ( *sp == 'm' )
       return(MVAL);
     else if ( *sp == 0 )
       return(NO);
     else
       mvpastFld();
   }
 }

/***/ fndFld(prefix)
 /* Moves ptr sp to start of first field starting with the chrs in
   the string, prefix.
   Returns: yes if found, no if not found before end of string.
*/
 char *prefix;
 {
   int result;
   do
     {
       while ( *sp == ' ' || *sp == '\t' )
         sp++;
```

```
        if ( (result = cmpPrefix(sp,prefix)) == YES )
          break;
      } while ( mvpastFld() );
     return( result );
  }

/***/ mvpastFld()
 /* Moves ptr sp past non-space chrs
    Returns: yes if done, no if end of line.
 */
 {
    while ( *sp != ' ' && *sp != '\0' && *sp != '\t' )
      sp++;
    return(*sp == '\0' ? NO : YES );
  }

/***/ cmpPrefix(sp,prefix)
 /* Compares start of string, sp, with chrs in prefix.
    Returns: yes if matched, no if not or end of string.
 */
  char *sp,
       *prefix;
 {
    if (*sp)
     {
       while ( *prefix == *sp )
         { prefix++; sp++; }
       return( *prefix == 0 );
     }
    else
      return(NO);
  }

/* end of file */
```

8a.2 Programme 2: Prediction of Reactor Operating Temperature - Phase I

Programme 2 predicts the operating temperature of the aerobic reactor for a given set of operating conditions (listed below) whilst oxygenation is with air alone. The oxygenation and heat balance terms were calibrated using the data derived from phase I and can be applied to both foaming and non-foaming conditions. The effect of incorporating pure oxygen supplementation can be estimated if required. The programme can be considered to embrace each of the models formulated in Chapter 5. Details regarding the compilation of the programme and the command line instruction are given in the programme. Appropriate operating data, encompassing all the main variables (listed below) are input from the command line after prompting from the programme.

$T(SL)_{in}$	The influent sludge temperature (°C)
$Q(SL)_{in}$	The influent sludge flow rate (m^3/d)
$Q(AIR)_{in}$	The influent air stream flow rate (m^3(STP)/h)
$M(O_2)_{sup}$	The pure oxygen supplementation rate, if required (m^3(STP)/h)
T_{amb}	The ambient temperature (°C)
$T(SL)_d$	The anaerobic digester sludge temperature, if known (°C)

Solution for the reactor sludge temperature $T(SL)_r$ is arrived at iteratively. The programme starts at a temperature of 40°C and increments by 0.01°C until the heat output from the system exceeds the heat input. The equations employed in the programme were derived in Chapters 3 and 5, in particular, the ***OTR*** functions in terms of ***OSR*** for foaming and non-foaming conditions (Equ's 5.28 and 5.29).

The accuracy of the model is demonstrated by simulating the appropriate operating data from the eight steady state periods during phase I (Table 4.2). The actual recorded temperatures and predicted temperatures along with the percentage errors are listed in Table 8a.1 below.

Table 8a.1 Comparison between Actual and Predicted Aerobic Reactor Sludge Temperatures for Each of the Steady State Periods During Phase I

Actual Temperature	Predicted Temperature	Percentage Error
55.9°	56.4°	+0.9
47.0°	47.4°	+0.9
49.9°	48.8°	-2.2
54.2°	54.3°	+0.2
43.7°	44.4°	+1.6
43.8°	44.0°	+0.5
48.5°	48.1°	-0.8
50.0°	49.6°	-0.8

Listing of Programme 2

```
/* Dual Digestion */
/*C Programme for the calculation of   */
/* reactor sludge temperature for a     */
/* specific set of operating conditions */
/* Compile Command: "cc -o progname -lm cprog.c "*/
/* Command Line: "progname"  */
/* Operating Data is Input From the Keyboard */
#include <stdio.h>
#include <ctype.h>
#include <math.h>
#define YES 1
#define NO 0
double sqrt();
double pow();
double OTR,OSR,Pvapo,Pvapi,TAIRo,TAIRi,Uo,Ui,Hg,Hs,TSLo,TSLi;
double Hv,Hw,Hm,Hb,Hi,Ho,Yh,YCO2,MOi;
double QSLi,Tamb,TSLd,QAIRi,O2,Ipump;
char Foam,Dig,Sup;
main()
{
 inputdata();
  calcTemp();
}
```

```
/***/inputdata()
{
 /*Reads in data from the Keyboard*/
 printf("              ********************************************\n");
 printf("              * CALCULATION OF REACTOR SLUDGE TEMPERATURE *\n");
 printf("              ********************************************\n");
 printf("\n");
 printf("\n");
 printf("Enter the Influent Sludge Flow Rate (m3/d)         : ");
 scanf("%lf",&QSLi);
 printf("Enter the Influent Air Flow Rate (m3/h)            : ");
 scanf("%lf",&QAIRi);
 printf("Enter the Ambient Temperature (°C)                 : ");
 scanf("%lf",&Tamb);
 printf("Is the Reactor Foaming (y/n)                       : ");
 scanf("%s",&Foam);
 printf("Do You Know The Digester Sludge Temperature (y/n)  : ");
 scanf("%s",&Dig);
  if (Dig == 'y')
    {
    printf("Enter Temperature (°C)                          : ");
    scanf("%lf",&TSLd);
    }
 printf("Is Pure Oxygen Supplementation Taking Place (y/n)  : ");
 scanf("%s",&Sup);
  if (Sup == 'y')
    {
    printf("Enter Pure Oxygen Supply Rate (kgO/h)           : ");
    scanf("%lf",&MOi);
    }
 printf("\n");
 printf("\n");
}
/***/calcTemp()
{
/* HEAT BALANCE - DUAL DIGESTER */
 TSLo  = 40;
  do
  {
  TSLo = TSLo + 0.01;
  /* CALCULATION OF THE BIOLOGICAL HEATING RATE */
  Yh    = 12.80;
  OSR   = 0.00151*QAIRi;
    if (Foam == 'y')
      {
      OTR = OSR*0.430*(exp(-0.822*OSR));
      }
    else
      {
      OTR = OSR*0.181*(exp(-0.321*OSR));
      }
```

```
  if (Sup == 'y')
    {
    Hb = (Yh*OTR*184.00)+(MOi*0.8*Yh);
    }
  else
    {
    Hb = Yh*OTR*184.00;
    }
/* CALCULATION OF MECHANICAL HEAT INPUT */
Ipump = 19;
Hm    = 1.65*Ipump;
/* CALCULATION OF OVERALL HEAT INPUT */
Hi    = Hb + Hm;
/* CALCULATION OF THE VAPOUR HEAT LOSS IN THE EFFLUENT GAS*/
TAIRi = Tamb;
  if (Foam == 'y')
    {
    TAIRo = TSLo;
    }
  else
    {
    TAIRo = TSLo-3.000;
    }
Pvapo = pow(10,(8.903-(2239.0/(273.0+TAIRo))));
Pvapi = pow(10,(9.124-(2307.0/(273.0+TAIRi))));
Uo    = 18.00*Pvapo/(29.00*(760.0-Pvapo));
Ui    = 18.00*Pvapi/(29.00*(1240.0-Pvapi));
Hv    = 2.382*1.205*QAIRi*(Uo-Ui);
/* CALCULATION OF SENSIBLE HEAT LOSS IN THE EFFLUENT GAS */
Hg    = ((1.205*QAIRi)/1000.00)*((1.00*(TAIRo-TAIRi))+(Uo*1.870*TAIRo));
/* CALCULATION OF SENSIBLE HEAT LOSS IN THE SLUDGE */
TSLi  = Tamb;
Hs    = (4.00/24.00)*((TSLo * QSLi)-((1.205*TSLo*QAIRi*Uo*24.00)/1000.00)-(QSLi*TSLi));
/* CALCULATION OF WALL HEAT LOSS */
  if (Dig == 'n')
    {
    TSLd  = ((TSLo+TSLi)/2.000);
    }
Hw    = (0.820*(TSLo-TSLd))+(0.490*(TSLo-TAIRi));
/* CALCULATION OF THE OVERALL HEAT OUTPUT */
Ho = Hv + Hs + Hw + Hg;
}
  while(Hi > Ho);
  printf("          ********************************************** \n");
  printf("          * The Predicted Reactor Temperature is %4.1lf * \n",TSLo);
  printf("          **********************************************\n");
}
```

8a.3 Programme 3: The General Dual Digestion Simulation Model

As a result of the investigation into the dual digestion system at Athlone (both phase I and phase II), sufficient information has been obtained with regard to the dual digestion system to allow a mathematical model to be compiled which can simulate all the main operating parameters in the system, and provide an assessment of system performance, both for the aerobic reactor and anaerobic digester, under a variety of different system configurations..

The model is able to simulate aerobic reactor performance using air or pure oxygen alone or a combination of air and pure oxygen. The model has been constructed in such a way that it can be applied for any size and type of dual digester application, i.e. it is not site specific for Athlone. The principal parameters which are computed by the model are as follows:

- Oxygenation characteristics of the aerobic reactor
- Steady state heat balance for the aerobic reactor
- Heating requirements for the anaerobic digester
- Effect of installing heat exchangers
- Volatile solids destruction in both the reactor and digester.
- Biogas production in the digester
- Performance of an installed gas engine
- The stability of the final sludge.
- Minimum reactor retention time to prevent substrate limitation.
- Minimum digester retention time to ensure sludge stability.

The model also includes a general cost analysis for both operating and capital costs. This allows cost comparison between different system configurations to be made. The model is also able to simulate conventional mesophilic anaerobic digestion and thermophilic digestion for comparison purposes.

The model has been compiled in the C programming language and can be run on either the UNIX or DOS operating system. The programme is started by running the batch file **ddsim**. Upon running the programme, the user is prompted to answer a number of questions in order to define the dual digestion system for simulation. Details regarding (1) the type of information required by the programme, (2) the computations performed by each component of the model and (3) the type of output produced is presented in Chapter 8.

Listing of Programme 3: The General Dual Digestion Simulation Model

```
/* DUAL DIGESTION: GENERAL SIMULATION PROGRAMME */
#include <stdio.h>
#include <ctype.h>
#include <math.h>
double sqrt();
double pow();
double QSLi,TSi,PVSi,PBVSi,PRBVSi,Vdig,RhDig,Vreac,RhReac;
double AreaDig,AreaReac,TSLd1,TSLd2,Tamb,TSLi,TSLr1,TSLr2,CpSL,Vp;
double QSLiDig,QSLiReac,UoDig,UoReac;
double Hx1trans,Hx2trans,Hx3trans,Hx4trans,Hx,Hx1eff,Hx2eff,Hx2effmech,Hx2power;
double PropHx2,PropHx3,Hx2max,Hx3eff,Hx4eff,AIReff,O2eff,MOut;
double CTSi,CVSi,CBVSi,CNBVSi,CRBVSi,CISi,MTSi,MVSi,MBVSi,MNBVSi,MRBVSi;
double CTSr,CVSr,CBVSr,CNBVSr,CISr,MTSr,MVSr,MBVSr,MNBVSr,VSremr,MBVSdr,CTSd;
double CVSd,CBVSd,CNBVSd,CISd,MTSd,MVSd,MBVSd,MNBVSd,VSremd,MBVSdd,MBVSdrmax,kd;
double bT,muT,FUNCTa,VSremtot,SOURest,CTSSd;
double MCH4gd,MCO2gd,MBGASgd,VBGASgd,GASprod,CCH4,HCH4,HCH4ex,HCH4ut,VCH4ex;
double VCH4ut,YCH4,fovs;
double Hcali,Hcalr,Hcald,Hsi,Hsr1,Hsr2,Hsd1,Hsd2,Hdig,Hreac,NoDigs,NoDigGAS;
double Hx1total,Hx2total,Hx3total,Hx4total,Hx1loss,Hx2loss,HlossHX3,Hx4loss;
double Hx2power,HCH4ex,EngReq,VEngReq,EngHrs;
double CapCost,OpCost,OxyCost,ElectCost,TransCost,RecovCost;
double TotCap,ReacCap,PumpCap,AirCap,DigCap,BoilCap,EngCap,ExchCap;
double OURair,OURo2,OUR,OURmax;
double Sa,Sa1,Sai,Zbh,Zbh1,Zbhi,Ksp,Kmp,Yzh,time,timemax,deltat;
double dSadt,dOdt,dZbhdt,bh,fe,SOUR,SOURmax;
double OTR,OSR,Pvapo,Pvapi,TAIRo,TAIRi,MAIRi,MGASi,Uo,Ui,Hg,Hs,TSLo,TSLi;
double Hv,Hw,Hm,Hb,Hi,Ho,Yh,YCO2,MOi;
double QAIRi,O2,Ipump,Oeff;
char   Option,OptShape,OptTemp,OptProg,OptSludge,OptAir,OptHX1,OptHX2;
char   OptHX3,OptHX4,OptWarn,Foam,Dig,Sup,page;

main()
{
  constants();
  inputdata(stderr);
  prog();
  energy();
  sourprog();
  costs();
  outdata();
}


/******PROGRAMME CONSTANTS*****/
/***/constants()
{
UoDig    = 0.00800;        /***Heat Transfer Coefficient MJ/°C.h.m2***/
UoReac   = 0.00733;        /***Heat Transfer Coefficient MJ/°C.h.m2***/
CpSL     = 4.0400;         /***Specific Heat Capacity of Sludge MJ/m3.°C***/
```

```
Hx1eff    = 0.6000;        /***Heat Transfer Efficiency of Interstage Cooler ***/
Hx2eff    = 0.4000;        /***Heat Transfer Efficiency of Exchanger on Gas Engine ***/
Hx2effmech= 0.3000;        /***Efficiency of Mechanical Energy Generation on Gas Engine***/
Hx3eff    = 0.6000;        /***Heat Transfer Efficiency of Exchanger on Conventional Heating ***/
Hx4eff    = 0.6000;        /***Transfer Efficiency of Afterstage Cooler ***/
Yh        = 12.823;        /***Specific Heat Yield MJ/h ***/
fovs      = 1.7000;        /***COD/VS Ratio ***/
YCH4      = 0.8820;        /**MJ/mol**/
}
/***END OF SUB-ROUTINE***/


/*********INPUT DATA SUB-ROUTINE************/
/***/inputdata(FILE *fp)
{
 fprintf(fp,"              ********************************************** \n");
 fprintf(fp,"              * GENERAL MODEL FOR THE DUAL DIGESTION PROCESS * \n");
 fprintf(fp,"              ********************************************** \n");
 fprintf(fp,"\n");
 fprintf(fp,"-------------------------------------\n");
 fprintf(fp,"Select Required Simulation Programme:\n");
 fprintf(fp,"-------------------------------------\n");
 fprintf(fp,"(a) User to Select System Retention Times\n");
 fprintf(fp,"(b) 1 day Reactor Retention Time & 10 day Digester Retention Time\n");
 fprintf(fp,"(c) 1 day Reactor Retention Time & Minimum Digester Retention Time\n");
 fprintf(fp,"(d) Minimum Reactor Retention Time & 10 day Digester Retention Time\n");
 fprintf(fp,"(e) Minimum Reactor Retention Time & Minimum Digester Retention Time\n");
 fprintf(fp,"(f) Conventional Anaerobic Digestion\n");
 fprintf(fp,"\n");
 fprintf(fp,"What is Your Response: ");
 scanf("%s",&OptProg);
   if (OptProg == 'a')
   {
   fprintf(fp,"Enter Required Aerobic Reactor Retention Time (days)    : ");
   scanf("%lf",&RhReac);
   fprintf(fp,"Enter Required Anaerobic Digester Retention Time (days) : ");
   scanf("%lf",&RhDig);
   }
   if (OptProg == 'f')
   {
   fprintf(fp,"Enter Required Anaerobic Digester Retention Time (days) : ");
   scanf("%lf",&RhDig);
   }
 fprintf(fp,"Enter the Size of the Digester(s) to be Upgraded/Constructed (m3)    : ");
 scanf("%lf",&Vdig);
 fprintf(fp,"What is the Total No of Streams to be Simulated(No):");
 scanf("%lf",&NoDigs);
 fprintf(fp,"What is the Shape of the Anaerobic Digester(s)?\n");
 fprintf(fp,"   (a) Anglo-Saxon Contour\n");
 fprintf(fp,"   (b) Classic Continental-European Contour\n");
 fprintf(fp,"   (c) Egg Shaped\n");
```

```
fprintf(fp,"\n");
fprintf(fp,"What is Your Response: ");
scanf("%s",&OptShape);
  if (OptShape == 'b')
     {
     AreaDig = 8.000*3.1415926*pow((Vdig/(3.000*3.1415926)),(2.000/3.000));
     }
  else if (OptShape == 'c')
     {
     AreaDig = 5.000*3.1415926*pow((Vdig/(2.000*3.1415926)),(2.000/3.000));
     }
  else
     {
     AreaDig = (8.000/3.000)*3.1415926*pow(((3.000*Vdig)/3.1415926),(2.000/3.000));
     }
fprintf(fp,"\n");
fprintf(fp,"Select Digester Temperature Range?\n");
fprintf(fp,"\n");
fprintf(fp,"  (a) Mesophilic (35°C)\n");
fprintf(fp,"  (b) Thermophilic (53°C)\n");
fprintf(fp,"  (c) Not Fixed (°C) Not Available for Conventional Digestion\n");
fprintf(fp,"\n");
fprintf(fp,"What is Your Response: ");
scanf("%s",&OptTemp);
fprintf(fp,"\n");
if (OptProg == 'f')
/***Options for Conventional Digestion***/
{
 fprintf(fp,"Do You Wish to Run a Gas Engine (y/n):");
 scanf("%s",&OptHX2);
 if (OptHX2 == 'y')
          {
          fprintf(fp,"What is the Power Rating of the Gas Engine (kW)? (Athlone = 1300kW):");
          scanf("%lf",&EngReq);
          EngReq = EngReq * 3.600;
          fprintf(fp,"Proportion of External Heat Required (percent):");
          scanf("%lf",&PropHx2);
          }
    else
          {
           fprintf(fp,"Do You Wish to Use Biogas to Heat Feed Sludge (y/n):");
           scanf("%s",&OptHX3);
           if (OptHX3 == 'y')
                {
                fprintf(fp,"Proportion of External Heat Required (percent):");
                scanf("%lf",&PropHx3);
                }
```

```
            else
                {
                fprintf(fp,"Do You Wish to Recover Heat From Effluent Digester Sludge(y/n):");
                scanf("%s",&OptHX4);
                }
            }
    }
  else
    {
/***Options for Dual Digestion***/
    fprintf(fp,"\n");
    fprintf(fp,"Do you wish to Supply Air to the Aerobic Reactor (y/n):");
    scanf("%s",&OptAir);
        if (OptAir == 'y')
            {
            fprintf(fp,"\n");
            fprintf(fp,"Enter the Air Flow Rate m3(STP)/h:");
            scanf("%lf",&QAIRi);
            fprintf(fp,"Enter the Anticipated Transfer Efficiency:");
            scanf("%lf",&AIReff);
            }
        if (OptTemp == 'c')
            {
            fprintf(fp,"\n");
            fprintf(fp,"Enter the Desired Aerobic Reactor Temperature (°C):");
            scanf("%lf",&TSLr1);
            }
       if (OptTemp == 'a')
            {
            fprintf(fp,"\n");
            fprintf(fp,"Do You Wish to Install Interstage Cooling (y/n):");
            scanf("%s",&OptHX1);
            }

        fprintf(fp,"Do You Wish to Run a Gas Engine (y/n):");
        scanf("%s",&OptHX2);
        if (OptHX2 == 'y')
            {
            fprintf(fp,"What is the Power Rating of the Gas Engine (kW)? (Athlone = 1300kW):");
            scanf("%lf",&EngReq);
            EngReq = EngReq*3.6000;
            fprintf(fp,"Proportion of External Heat Required (percent):");
            scanf("%lf",&PropHx2);
            }
        else
            {
             fprintf(fp,"Do You Wish to Use Biogas to Heat Feed Sludge (y/n):");
             scanf("%s",&OptHX3);
```

```
            if (OptHX3 == 'y')
                {
                fprintf(fp,"Proportion of External Heat Required (percent):");
                scanf("%lf",&PropHx3);
                }
            else
                {
                fprintf(fp,"Do You Wish to Recover Heat From Effluent Digester Sludge(y/n):");
                scanf("%s",&OptHX4);
                }
            }
        }
fprintf(fp,"\n");
fprintf(fp,"FEED SLUDGE CHARACTERISTICS\n");
fprintf(fp,"--------------------------\n");
fprintf(fp,"What Type of Sludge in being Treated?\n");
fprintf(fp,"   (a) Primary Sludge\n");
fprintf(fp,"   (b) Primary/Humus Sludge Mixture\n");
fprintf(fp,"   (c) Primary/Waste Activated Sludge Mixture\n");
fprintf(fp,"   (d) Waste Activated Sludge\n");
fprintf(fp,"   (e) User to Select Sludge Characteristics\n");
fprintf(fp,"\n");
fprintf(fp,"What is Your Response: ");
scanf("%s",&OptSludge);
fprintf(fp,"\n");
fprintf(fp,"Enter the Expected Feed Sludge Total Solids Concentration (kg/m3)  : ");
scanf("%lf",&TSi);
  if (OptSludge == 'a')
      {
      PVSi   =  81.000;
      PBVSi  =  60.000;
      PRBVSi =  20.000;
      }
  if (OptSludge == 'b')
      {
      PVSi   =  78.000;
      PBVSi  =  50.000;
      PRBVSi =  10.000;
      }
  if (OptSludge == 'c')
      {
      PVSi   =  75.000;
      PBVSi  =  45.000;
      PRBVSi =  10.000;
      }
  if (OptSludge == 'd')
      {
      PVSi   =  70.000;
      PBVSi  =  40.000;
      PRBVSi =  0.5000;
      }
```

```
   if (OptSludge == 'e')
      {
 fprintf(fp,"Enter the Volatile Solids Fraction of the Total Solids (percent)   : ");
 scanf("%lf",&PVSi);
 fprintf(fp,"Enter the Biodegradable Fraction of the Volatile Solids (percent)  : ");
 scanf("%lf",&PBVSi);
 fprintf(fp,"Enter the Readily Biodeg Fraction of the Biodeg VS (percent)       : ");
 scanf("%lf",&PRBVSi);
      }
 fprintf(fp,"Enter the Ambient Temperature for Modelling (°C)                   : ");
 scanf("%lf",&Tamb);
 fprintf(fp,"\n");
 fprintf(fp,"\n");
 fprintf(fp,"\n");
 fprintf(fp,"\n");
 fprintf(fp,"\n");
 fprintf(fp,"\n");
 fprintf(fp,"\n");
 fprintf(fp,"\n");
}
/***END OF SUB-ROUTINE***/


/***PROGRAMME SELECTION SUB-ROUTINE***/
/***/prog()
{
 if (OptProg == 'a')
 /** User has Selected Retention Times**/
   {
   QSLi  = Vdig/RhDig;
   Vreac = QSLi * RhReac;
   heatcalcs();
   }
 else if (OptProg == 'b')
 /** 1 day Reactor Retention Time & 10 day Digester Retention Time**/
   {
   RhDig =10.000;
   RhReac=1.000;
   QSLi  = Vdig/RhDig;
   Vreac = QSLi * RhReac;
   heatcalcs();
   }
 else if (OptProg == 'c')
 /** 1 day Reactor Retention Time & Minimum Digester Retention Time**/
   {
   RhDig  =9.900;
   RhReac =1.000;
      do
      {
      RhDig  = RhDig + 0.100;
      QSLi  = Vdig/RhDig;
```

```
        Vreac = QSLi * RhReac;
        heatcalcs();
        }
        while (CBVSd >= 2.500 );
    }
 else if (OptProg == 'd')
/** Minimum Reactor Retention Time & 10 day Digester Retention Time**/
    {
    RhReac = 4.000;
    RhDig  = 10.000;
    QSLi  = Vdig/10.000;
    do
      {
      RhReac = RhReac - 0.010;
      Vreac = QSLi * RhReac;
      heatcalcs();
        if (OUR > OURmax)
        {
        return;
        }
      }
    while (RhReac >= 0.900 );
    }
 else if (OptProg == 'e')
/** Minimum Reactor Retention Time & Minimum Digester Retention Time**/
    {
    RhReac = 4.000;
    RhDig  = 10.000;
    QSLi  = Vdig/10.000;
    do
      {
      RhReac = RhReac - 0.010;
      Vreac = QSLi * RhReac;
      heatcalcs();
        if (OUR > OURmax)
        {
        return;
        }
      }
    while (RhReac >= 0.900 );
        do
        {
        RhDig  = RhDig + 0.100;
        QSLi  = Vdig/RhDig;
        Vreac = QSLi * RhReac;
        heatcalcs();
        }
        while (CBVSd >= 2.500 );
    }

 else if (OptProg == 'f')
```

```
/** Conventional Anaerobic Digestion**/
   {
  QSLi  = Vdig/RhDig;
  Vreac = 0;
  CTSi  = TSi;
  CVSi  = TSi*(PVSi/100.000);
  CISi  = CTSi-CVSi;
  CBVSi = TSi*(PVSi/100.000)*(PBVSi/100.000);
  CRBVSi= TSi*(PVSi/100.000)*(PBVSi/100.000)*(PRBVSi/100.000);
  CNBVSi= CVSi-CBVSi;
  TAIRi = Tamb;
  MTSi  = (QSLi/24.000)*TSi;
  MVSi  = (QSLi/24.000)*TSi*(PVSi/100.000);
  MBVSi = (QSLi/24.000)*TSi*(PVSi/100.000)*(PBVSi/100.000);
  MNBVSi= MVSi-MBVSi;
  if (OptTemp == 'b')
   {
  TSLd1 = 53.00000;
   }
  else
   {
  TSLd1 = 35.00000;
   }
  TSLr2 =  (((24.000*UoDig*AreaDig)/(CpSL*QSLi))*(TSLd1 - Tamb))+TSLd1;
  TSLr1 =TSLr2;
  MOi   = 0.00;
  MOut  = 0.00;
  Hb = 0.00;
  Hm  = 0.00;
  Hv    = 0.000;
  Hg    = 0.000;
  TSLi  = Tamb;
  Hs    = (4.04/24.00)*QSLi*(TSLr1 - TSLi);
  AreaReac = 8.000*3.1415926*pow((Vreac/(3.000*3.1415926)),(2.000/3.000));
  Ho = Hv + Hs + Hw + Hg;
  vscalcs();
  Hx = Hx1trans + Hx2trans + Hx3trans +Hx4trans;
  Hi = Hb + Hm + Hx;
  if (Hi < Ho + 1.000)
   {
   OptWarn = 'y';/**Warning Signal for Conventional Digestion- Insufficient Heat**/
   }
  }
}
/***END OF SUB-ROUTINE***/
```

```
/* HEAT BALANCE - SUB-ROUTINE*/
/***/heatcalcs()
{
  CTSi  = TSi;
  CVSi  = TSi*(PVSi/100.000);
  CISi  = CTSi-CVSi;
  CBVSi = CVSi*(PBVSi/100.000);
  CRBVSi = CVSi*(PRBVSi/100.000);
  CNBVSi= CVSi-CBVSi;
  MTSi  = (QSLi/24.000)*TSi;
  MVSi  = MTSi*(PVSi/100.000);
  MBVSi = MVSi*(PBVSi/100.000);
  MRBVSi = MBVSi*(PRBVSi/100.000);
  MNBVSi= MVSi-MBVSi;
  TAIRi = Tamb;
/***Calculation of System Temperatures***/
    if (OptTemp == 'a')
       {
    TSLd1 = 35.00000;
    TSLr2 =  (((24.000*UoDig*AreaDig)/(CpSL*QSLi))*(TSLd1 - Tamb))+TSLd1;
          if (OptHX1 == 'y')
            {
            TSLr1 = 60.000;
            }
          else
            {
            TSLr1 =TSLr2;
            }
       }
    else if (OptTemp == 'c')
       {
    TSLr2 = TSLr1;
    TSLd1 =
((((TSLr2*CpSL*QSLi)/(24.000*UoDig*AreaDig))+Tamb)/(1.000+((CpSL*QSLi)/(24.000*UoDig*AreaDig))));
       }
    else
       {
    TSLd1 = 53.0000;
    TSLr2 =  (((24.000*UoDig*AreaDig)/(CpSL*QSLi))*(TSLd1 - Tamb))+TSLd1;
    TSLr1 = TSLr2;
       }

MAIRi  = 1.205 * QAIRi;
Oeff   = 80.000;

 MOi   = 0.0000;
  do
  {
  /* CALCULATION OF THE BIOLOGICAL HEATING RATE */
  MOut = ((MAIRi*0.230*AIReff*0.0100)+(MOi*Oeff*0.0100));
  OUR    = MOut/Vreac;
```

```
  OURair = MAIRi*0.230*AIReff*0.0100/Vreac;
  OURo2  = MOi*Oeff*0.0100/Vreac;
  Hb = Yh*MOut;
  /* CALCULATION OF MECHANICAL HEAT INPUT */
Ipump = Vreac;
Hm    = 1.65 * Ipump;
  /* HEAT RECOVERY */
  Hx = Hx1trans + Hx2trans + Hx3trans +Hx4trans;
  /* CALCULATION OF OVERALL HEAT INPUT */
  Hi = Hb + Hm + Hx;
  /* CALCULATION OF THE VAPOUR HEAT LOSS IN THE EFFLUENT GAS*/
  TAIRo = TSLr1 -3.000;
  Pvapo = pow(10,(8.903-(2239.0/(273.0+TAIRo))));
  Pvapi = pow(10,(9.124-(2307.0/(273.0+TAIRi))));
  Uo    = 18.00*Pvapo/(29.00*(760.0-Pvapo));
  Ui    = 18.00*Pvapi/(29.00*(1240.0-Pvapi));
  MGASi = MAIRi + MOi;
  Hv    = 2.382*MGASi*(Uo-Ui);
  /* CALCULATION OF SENSIBLE HEAT LOSS IN THE EFFLUENT GAS */
  Hg    = ((MGASi)/1000.00)*((1.00*(TAIRo-TAIRi))+(Uo*1.870*TAIRo));
  /* CALCULATION OF SENSIBLE HEAT LOSS IN THE SLUDGE */
  TSLi  = Tamb;
  Hs    = (4.00/24.00)*((TSLr1 * QSLi)-((MGASi*TSLr1*Uo*24.00)/1000.00)-(QSLi*TSLi));
  /* CALCULATION OF WALL HEAT LOSS */
  /**based on h=3r***/
  AreaReac = 8.000*3.1415926*pow((Vreac/(3.000*3.1415926)),(2.000/3.000));
  Hw = UoReac * AreaReac * (TSLr1 - Tamb);
  /* CALCULATION OF THE OVERALL HEAT OUTPUT */
  Ho = Hv + Hs + Hw + Hg;
  vscalcs();
  if (MOi <= 0.2 && Hi >= Ho)
   {
   MOi = 0.000;
   Hb  = Yh*(MAIRi*0.230*AIReff*0.0100);
   return;
   }
  else
   {
  MOi  = MOi + 0.1;
   }
}
  while(Hi < Ho);
}
/***END OF SUB-ROUTINE***/
```

```
/****VS DESTRUCTION SUB-ROUTINE****/
/***/vscalcs()
{
kd     = (0.0025 * TSLr1) + 0.125;
MBVSdrmax = ((kd*Vreac*CBVSi)/(24.000*(1.000+(kd*RhReac))))+((Vreac*CRBVSi)/(24.000*RhReac));
CISr   = CISi;
CNBVSr = CNBVSi;
MBVSdr = MOut/fovs;
MBVSr  = MBVSi-MBVSdr;
CBVSr  = MBVSr / (QSLi/24.000);
CVSr   = CBVSr+CNBVSr;
CTSr   = CISr+CVSr;
VSremr = 100.000*((CVSi-CVSr)/CVSi);
OURmax = fovs * MBVSdrmax/Vreac;
/****VS DESTRUCTION IN DIGESTER****/
CISd   = CISr;
bT     = 0.020 * pow(2.71828,(0.0600 * (TSLd1 - 35.000)));
muT    = 0.324 * pow(2.71828,(0.1400 * (TSLd1 - 35.000)));
FUNCTa = (1 + (bT * RhDig))/(muT * RhDig);
CBVSd  = 0.597*FUNCTa*CBVSr;
CVSd   = CNBVSr + (0.177*CBVSr) + (0.582*FUNCTa*CBVSr);
CNBVSd = CVSd - CBVSd;
CTSd   = CISd + CVSd;
MBVSdd = (QSLi/24.000)*(CVSr-CVSd);
VSremd = 100.000*((CVSr-CVSd)/CVSr);
VSremtot= 100.000*((CVSi-CVSd)/CVSi);
MBVSd  = CBVSd *(QSLi/24.0000);
CTSSd  = CTSd - CBVSd;
SOURest    = fovs * ((1000.000*CBVSd) / CTSSd);
/****BIOGAS PRODUCTION IN THE DIGESTER****/
/***UNITS mol/h***/
MCH4gd = fovs * CBVSr * (QSLi/24.000) * (12.830 - (9.080*FUNCTa));
MCO2gd = fovs * CBVSr * (QSLi/24.000) * (7.690  + (0.104*FUNCTa));
MBGASgd= fovs * CBVSr * (QSLi/24.000) * (20.520 - (8.980*FUNCTa));
VBGASgd= 0.0224*MBGASgd;
CCH4   = 100.00 * MCH4gd / MBGASgd;
HCH4   = YCH4 * MCH4gd;
GASprod = VBGASgd / MBVSdd;
heatx();
}
/***END OF SUB-ROUTINE***/


/***HEAT EXCHANGER SUB-ROUTINE****/
/***/heatx()
{
   /**INTERSTAGE COOLER**/
   if (OptHX1 == 'y')
     {
     Hx1total = QSLi * CpSL * (60.000 - TSLr2)/24.000;
     Hx1trans = Hx1eff * Hx1total;
```

```
  Hx1loss = Hx1total - Hx1trans;
  }
/**GAS ENGINE**/
if (OptHX2 == 'y')
  {
  Hx2max = EngReq/NoDigs;
      if (HCH4 >= Hx2max)
      {
      Hx2total = Hx2max;
      }
      else
      {
      Hx2total = HCH4;
      }
      Hx2trans = Hx2eff * Hx2total;
      Hx2power = Hx2effmech * Hx2total;
      EngHrs = (24.000*Hx2total)/EngReq;
      Hx2loss = Hx2total - Hx2trans - Hx2power;
          if (Hx2trans > (PropHx2/100.000)*(Ho-Hx1trans-Hx4trans-Hm) )
              {
              Hx2trans = (PropHx2/100.000)*(Ho - Hx1trans - Hx4trans -Hm);
              Hx2loss = Hx2total - Hx2trans -Hx2power;
              }
  }
/**CONVENTIONAL BOILER**/
if (OptHX3 == 'y')
  {
  Hx3total = HCH4;
  Hx3trans = Hx3eff * Hx3total;
  HlossHX3 = Hx3total - Hx3trans;
       if (Hx3trans > (PropHx3/100.000)*(Ho-Hx1trans-Hx4trans-Hm) )
              {
              Hx3trans = (PropHx3/100.000)*(Ho - Hx1trans - Hx4trans -Hm);
              Hx3total = Hx3trans /Hx3eff;
              HlossHX3 = Hx3total - Hx3trans;
              }
       else
              {
  Hx2power = Hx2effmech * HCH4;
  EngHrs = (24.000*HCH4)/EngReq;
              }
  }
/**AFTERSTAGE COOLER**/
if (OptHX4 == 'y')
  {
  TSLd2 = (0.625*Tamb) + (0.375*TSLd1);
  Hx4total = (QSLi/24.000) * CpSL * (TSLd1 - TSLd2);
  Hx4trans = Hx4eff * Hx4total;
  Hx4loss = Hx4total - Hx4trans;
```

```
            if (Hx4trans > Ho-Hx2trans-Hx3trans-Hm )
                    {
                    Hx4trans = Ho - Hx2trans -Hx3trans - Hm;
                    TSLd2 = TSLd1 - (Hx4trans/Hx4eff)*(24.000/(CpSL*QSLi));
                    Hx4total = (QSLi/24.000)*CpSL*(TSLd1 - TSLd2);
                    Hx4loss = Hx4total - Hx4trans;
                    }
      }
     else
      {
       TSLd2 = TSLd1;
      }
}
/***END OF SUB-ROUTINE***/


/****SYSTEM ENERGIES SUB-ROUTINE****/
/***/energy()
{
/***Calorific Values***/
Hcali   = Yh * fovs * MBVSi;
Hcalr   = Hcali - Hb;
Hcald   = Hcali - HCH4 - Hb;
/***Sludge Sensible Heats***/
Hsi     = 0.000;
Hsr1    = (4.04/24.00)*QSLi*(TSLr1 - Tamb);
Hsr2    = (4.04/24.00)*QSLi*(TSLr2 - Tamb);
Hsd1    = (4.04/24.00)*QSLi*(TSLd1 - Tamb);
Hsd2    = (4.04/24.00)*QSLi*(TSLd2 - Tamb);
/***Used and Unused Biogas Energy***/
VEngReq  = ((24.000*100.000*0.0224*EngReq)/(CCH4*YCH4));
HCH4ut   = Hx2total + Hx3total;
HCH4ex   = HCH4 - HCH4ut;
VCH4ut  =  NoDigs*24.000*(100.000/CCH4)*0.0224 * HCH4ut / YCH4;
VCH4ex  =  NoDigs*24.000*(100.000/CCH4)*0.0224 * HCH4ex / YCH4;
/***Heat Losses From Reactor and Digester***/
Hreac   =  Hv + Hg + Hw;
Hdig    =  (UoDig*AreaDig)*(TSLd1 - Tamb);
}
/***END OF SUB-ROUTINE***/


/***SOUR TEST SUB-ROUTINE***/
/***/sourprog()
{
Sai    =  fovs*(CBVSd);
Kmp = 5.5000/24.000;    /***kgCOD/kgCOD-cells.h***/
Yzh = 0.660;            /***kgCOD/kgCOD-cells***/
Ksp = 0.027;            /***kgCOD/kgCOD***/
Zbhi = 0.25;            /***kgCOD/m3****/
deltat = 0.01;          /****h****/
```

```
fe   = 0.200;            /***kgCOD/kgCOD***/
bh   = 0.01;             /*** /h***/
Zbh = Zbhi;
Sa  = Sai;
time = 0.000;
do
{
  dSadt  =  (-1.000/Yzh)*Kmp*((Sa/Zbh)/(Ksp+(Sa/Zbh)))*Zbh + (1.000-fe)*bh*Zbh;
  dZbhdt  =  (Yzh*(1.000/Yzh)*Kmp*((Sa/Zbh)/(Ksp+(Sa/Zbh)))*Zbh) - (bh*Zbh);
  Sa1   = Sa + (dSadt*deltat);
  Sa    = Sa1;
  Zbh1   = Zbh + (dZbhdt*deltat);
  Zbh    = Zbh1;
  dOdt   =  (1000.000*(1.000-Yzh)/Yzh)*Kmp*((Sa/Zbh)/(Ksp+(Sa/Zbh)))*Zbh;
  if ((dOdt/CTSd) > SOUR)
    {
    SOURmax = SOUR;
    timemax = time;
    }
  SOUR   =  dOdt/CTSd;
  time   = time + deltat;
}
while (time <= 96.00);
}
/***END OF SUB-ROUTINE***/


/****PRINTING SUBROUTINE****/
/***/outdata()
{
char buff[80];
dispdata(stderr);
fprintf(stderr,"Do You Wish to Print? (y/n):");
scanf("%s",buff);
if (buff[0] == 'y' )
      {
      dispdata(stdout);
      }
}
/***END OF SUB-ROUTINE***/


/****OPTIONS FOR PRINTING DATA****/
/***/dispdata(FILE *fp)
{
fprintf(fp,"\n");
fprintf(fp,"SYSTEM DATA: DUAL DIGESTION PLANT\n");
fprintf(fp,"Feed Sludge Flow Rate of");
```

```
    if (OptSludge == 'a')
      {
fprintf(fp," Primary Sludge");
      }
    if (OptSludge == 'b')
      {
fprintf(fp," Primary Sludge/Humus Sludge Mixture");
      }
    if (OptSludge == 'c')
      {
fprintf(fp," Primary Sludge/Waste Activated Sludge Mixture");
      }
    if (OptSludge == 'd')
      {
fprintf(fp," Waste Activated Sludge");
      }
    if (OptSludge == 'e')
      {
fprintf(fp," User Defined Sludge");
      }
fprintf(fp," %4.0lf m3/d\n",NoDigs*QSLi);
fprintf(fp,"Number of Streams     %4.0lf      Ambient Temperature      %5.1lf C \n",NoDigs,Tamb);
fprintf(fp,"Oxygenation of Aerobic Reactor: ");
    if (MOi > 0.000)
      {
fprintf(fp,"Pure Oxygen ");
      }
    if (OptAir == 'y')
      {
fprintf(fp,"Air ");
      }
    if (OptAir != 'y' && MOi < 0.100)
      {
fprintf(fp,"None. Equivalent to Conventional Digestion ");
      }
fprintf(fp,"\n");
fprintf(fp,"Aerobic Reactor Size  %4.0lf m3   Anaerobic Digester Size  %5.0lf m3 \n",Vreac,Vdig);
fprintf(fp,"Retention Times:  Reactor  %5.2lf days     Digester %5.2lf days    \n",RhReac,RhDig);
fprintf(fp,"Ext Heating Units: ");
    if (OptHX1 == 'y')
      {
fprintf(fp,"Interstage Heat Exchange: ");
      }
    if (OptHX2 == 'y')
      {
fprintf(fp,"Gas Engine: ");
      }
    if (OptHX3 == 'y')
      {
fprintf(fp,"Boilers: ");
      }
```

```
    if (OptHX4 == 'y')
      {
fprintf(fp,"Afterstage Heat Exchange: ");
      }
    if (OptHX1 == 'y' || OptHX2 == 'y' || OptHX3 == 'y' || OptHX4 == 'y')
      {
fprintf(fp,"\n");
      }
    else
      {
fprintf(fp,"No External Heating, Gas Engine, Boilers etc Provided.\n");
      }
fprintf(fp,"\n");
fprintf(fp,"SLUDGE QUALITY  INFLUENT SLUDGE   REACTOR SLUDGE   DIGESTER SLUDGE\n");
fprintf(fp,"TS(kg/m3)          %6.1lf          %6.1lf          %6.1lf     \n",CTSi,CTSr,CTSd);
fprintf(fp,"VS(kg/m3)          %6.1lf          %6.1lf          %6.1lf     \n",CVSi,CVSr,CVSd);
fprintf(fp,"BVS(kg/m3)         %6.1lf          %6.1lf          %6.1lf     \n",CBVSi,CBVSr,CBVSd);
fprintf(fp,"Percent VS (d.m.b.) %6.1lf          %6.1lf          %6.1lf
\n",100*CVSi/CTSi,100*CVSr/CTSr,100*CVSd/CTSd);
fprintf(fp,"Temperature (C)     %6.1lf          %6.1lf          %6.1lf     \n",Tamb,TSLr1,TSLd1);
fprintf(fp,"PERCENT          AEROBIC REACTOR ANAEROBIC DIGESTER   OVERALL\n");
fprintf(fp,"VS REMOVAL          %6.1lf          %6.1lf          %6.1lf     \n",VSremr,VSremd,VSremtot);
fprintf(fp,"\n");
fprintf(fp,"OUR (kg/m3.h) Oxygen = %5.3lf Air = %5.3lf Total = %5.3lf (maximum =
%5.3lf)\n",OURo2,OURair,OUR,OURmax);
fprintf(fp,"\n");
    if (NoDigs == 1)
    {
fprintf(fp,"OXYGEN CONSUMPTION:%6.1lf kg/h    %6.2lf T/d   %6.0lf
T/annum\n",MOi,0.024*MOi,0.024*365.000*MOi);
fprintf(fp,"\n");
fprintf(fp,"BIOGAS PRODUCTION:%6.1lf m3/h     %6.0lf m3/d \n",VBGASgd,24.000*VBGASgd);
    }
    else
    {
fprintf(fp,"OXYGEN CONSUMPTION: Per Reactor   = %6.1lf kg/h  %6.2lf T/d  %6.0lf
T/annum\n",MOi,0.024*MOi,0.024*365.000*MOi);
fprintf(fp,"                    Total         = %6.1lf kg/h  %6.2lf T/d  %6.0lf
T/annum\n",NoDigs*MOi,NoDigs*0.024*MOi,NoDigs*0.024*365.000*MOi);
fprintf(fp,"\n");
fprintf(fp,"BIOGAS PRODUCTION: Per Digester     = %6.1lf m3/h    %6.0lf m3/d
\n",VBGASgd,24.000*VBGASgd);
fprintf(fp,"                   Total For System  = %6.1lf m3/h    %6.0lf m3/d
\n",NoDigs*VBGASgd,NoDigs*24.000*VBGASgd);
    }
fprintf(fp,"                   Utilised  = %6.0lf m3/d   Wasted  = %6.0lf m3/d \n",VCH4ut,VCH4ex);
fprintf(fp,"BIOGAS QUALITY     Methane  = %5.1lf .     Carbon Dioxide = %5.1lf
percent\n",CCH4,100.000-CCH4);
fprintf(fp,"\n");
```

```
/* Next display page */
if (fp == stderr)
  {
    fprintf(stderr,"Do You Wish to Continue y/n:");
    scanf("%s",&page);
  }
if (page != 'y')
  {
   return(0);
  }
fprintf(fp,"ENERGY BALANCE AEROBIC REACTOR\n");
fprintf(fp,"Biological Heat     %6.0lf       Wall Heat Loss        %6.0lf\n",Hb,Hw);
fprintf(fp,"Mechanical Heat     %6.0lf       Vent Gas Heat Loss    %6.0lf\n",Hm,Hg+Hv);
fprintf(fp,"Recycled Heat       %6.0lf       Sensible Heat Loss    %6.0lf\n",Hx,Hs);
fprintf(fp,"Total Heat In       %6.0lf       Total Heat Out        %6.0lf\n",Hi,Ho);
fprintf(fp,"\n");
fprintf(fp,"ENERGY BALANCE ANAEROBIC DIGESTER\n");
fprintf(fp,"Sensible Heat Given %6.0lf       Wall Heat Loss        %6.0lf\n",Hsr2-Hsd1,Hdig);
fprintf(fp,"\n");
fprintf(fp,"HEAT EXCHANGERS\n");
    if (OptHX1 == 'y')
    {
fprintf(fp,"InterStage Heat Exchange: Total Heat Available %6.0lf = Transferred %6.0lf + Lost
%6.0lf\n",Hx1total,Hx1trans,Hx1loss);
fprintf(fp,"Sludge Temp in = %5.1lf  Sludge Temp Out = %5.1lf\n",TSLr1,TSLr2);
    }
    if (OptHX2 == 'y')
    {
fprintf(fp,"Gas Engine Heat: Total %6.0lf = Transf %6.0lf + Lost %6.0lf + Mech
%6.0lf\n",Hx2total,Hx2trans,Hx2loss,Hx2power);
    }
    if (OptHX3 == 'y')
    {
fprintf(fp,"Boilers: Total Biogas Energy Burnt %6.0lf = Transferred %6.0lf + Lost
%6.0lf\n",Hx3total,Hx3trans,HlossHX3);
    }
    if (OptHX4 == 'y')
    {
fprintf(fp,"Afterstage Heat Exchange: Total Heat Available %6.0lf = Transferred %6.0lf + Lost
%6.0lf\n",Hx4total,Hx4trans,Hx4loss);
fprintf(fp,"Sludge Temp in = %5.1lf  Sludge Temp Out = %5.1lf\n",TSLd1,TSLd2);
    }
fprintf(fp,"\n");
    if (OptHX2 == 'y')
    {
fprintf(fp,"    Gas Engine Requires             %6.1lf m3/d \n",VEngReq);
fprintf(fp,"    Gas Engine Running Time per Day %6.1lf Hours (%6.1lf
percent)\n",NoDigs*EngHrs,((100.000*NoDigs*EngHrs)/24.000));
    }
```

```
  if (OptWarn == 'y')
   {
fprintf(fp,"WARNING: INSUFFICIENT HEAT INPUT TO REACH RECOMMENDED DIGESTION
TEMPERATURE\n");
   }
fprintf(fp,"\n");
fprintf(fp,"SLUDGE CALORIFIC VALUES (Based in BVS Concentration)\n");
fprintf(fp,"Feed Sludge %6.0lf    Reactor Sludge %6.0lf    Digester Sludge %6.0lf\n",Hcali,Hcalr,Hcald);
fprintf(fp,"Calorific Value Reduced by:  Oxidation  %6.0lf    Biogas Production %6.0lf\n",Hb,HCH4);
fprintf(fp,"\n");
fprintf(fp,"SENSIBLE HEATS (In Relation to the Feed Sludge Temperature (%5.1lfC)\n",Tamb);
fprintf(fp,"Feed: %2.0lf Reactor: %4.0lf ",Hsi,Hsr1);
    if (OptHX1 == 'y')
    {
fprintf(fp,"After Heat Exchange: %4.0lf ",Hsr2);
    }
fprintf(fp,"Digester: %4.0lf ",Hsd1);
    if (OptHX4 == 'y')
    {
fprintf(fp,"After Heat Exchange: %4.0lf ",Hsd2);
    }
fprintf(fp,"\n");
fprintf(fp,"\n");
fprintf(fp,"FINAL SLUDGE QUALITY\n");
fprintf(fp,"Biodegradable Volatile Solids     %8.2lf kgBVS/m3      \n",CBVSd);
fprintf(fp,"Total Oxygen Demand               %8.3lf gO/kgTSS    \n",SOURest);
fprintf(fp,"The Predicted Maximum SOUR  %8.3lf gO/kgTSS.h\n",SOURmax);
fprintf(fp,"(Occurring after %6.1lf hrs of the test)\n",timemax);
fprintf(fp,"\n");
/* Next display page */
if (fp == stderr)
 {
   fprintf(stderr,"Do You Wish to Continue y/n:");
   scanf("%s",&page);
 }
if (page != 'y')
 {
  return(0);
 }
fprintf(fp,"CAPITAL COST (1995)   OPERATING COSTS P.A. (2002)   (R1,000,000's)\n");
fprintf(fp,"Reactor    %6.3lf     Oxygen         %6.3lf\n",ReacCap,OxyCost);
fprintf(fp,"Pumps      %6.3lf     Electrical     %6.3lf\n",PumpCap,ElectCost);
fprintf(fp,"Air Supply %6.3lf     Transport      %6.3lf\n",AirCap,TransCost);
fprintf(fp,"Digester   %6.3lf     Sub Total      %6.3lf\n",DigCap,OxyCost+ElectCost+TransCost);
fprintf(fp,"Boiler     %6.3lf     Recovered      %6.3lf\n",BoilCap,RecovCost);
fprintf(fp,"Gas Engine %6.3lf     Sub Total      %6.3lf\n",EngCap,OpCost);
fprintf(fp,"Exchangers %6.3lf     Cap Repayment  %6.3lf\n",ExchCap,CapCost);
fprintf(fp,"Total      %6.3lf     Total          %6.3lf\n",TotCap,OpCost+CapCost);
}
/***END OF SUB-ROUTINE***/
```

```
/****SUBROUTINE FOR COSTS****/
/***/costs()
{
ReacCap   = NoDigs*1.450*pow((0.001*Vreac),0.63);
PumpCap   = NoDigs*0.002*(0.460*Vreac);
AirCap    = NoDigs*(62.5*QAIRi)/1000000.000;
DigCap    = NoDigs*1.940*pow((0.001*Vdig),0.63);
if (OptHX3 == 'y')
 {
BoilCap   = NoDigs*0.5000*(VCH4ut/1000.000);
 }
else
 {
BoilCap   = 0.000;
 }
EngCap    = (EngReq/3.600)*(5.000/1400);
ExchCap   = NoDigs*0.000200*((Hx1trans+Hx2trans+Hx3trans+Hx4trans)/3.600);
TotCap    = ReacCap+PumpCap+AirCap+DigCap+BoilCap+EngCap+ExchCap;

CapCost   = 0.19325*TotCap;

OxyCost   = 2.85*NoDigs*(0.450 * MOi * 24.000 * 365.000)/1000000.000;
ElectCost =
2.85*NoDigs*(0.200*((0.460*Vreac)+(0.025*QAIRi)+(0.050*Vdig))*24.000*365.000)/1000000.000;
if (TSLr1 < 60.00 && TSLd1 < 53.000)
{
TransCost = 2.85*NoDigs*0.032*QSLi*CTSd*365.000/1000000.000;
}
else
{
TransCost = 0.0000;
}
RecovCost = 2.85*NoDigs*0.200*(Hx2power/3.600)*24.000*365.000/1000000;
OpCost    = OxyCost+ElectCost+TransCost-RecovCost;
}
/***END OF SUB-ROUTINE***/
```

APPENDIX 9

THE MOTIVATION FOR PHASE II AND THE PROPOSED PROGRAMME: PURE OXYGEN SUPPLEMENTATION

9a.1 Introduction

From the conclusions made in Chapter 6, it was recommended that further investigative work (phase II) be carried out on the Athlone Dual Digestion Plant, to evaluate the feasibility of supplementing the aeration of the aerobic reactor with pure oxygen. It was envisaged that with pure oxygen supplementation a substantial proportion of the primary sludge produced at Athlone could be pasteurised and stabilised by the process without adversely affecting biogas production, which is required as fuel for the gas engine.

The primary sludge produced at Athlone is currently treated by conventional anaerobic digestion. In terms of the guidelines produced by the Department of National Health and Population Development (DNH&PD, 1991; see Section 1.2.1) the treated sludge is classified as a type B sludge and is therefore unsuitable for agricultural use. At present, this sludge is transported after drying to the Coastal Park Refuse Disposal Site. The negative aspects of this current practice are; (1) the beneficial use of the sludge as a soil conditioner and source of nutrients is lost, (2) the life span of the Refuse Disposal Site is reduced, and (3) significant costs are involved in transporting the sludge to the Disposal Site.

The options available for the disposal of the sludge at Athlone would be greatly increased if that sludge were to be pasteurised. An increase in the monitoring of the incoming raw wastewater at Athlone has placed tighter control on the discharge of abnormally high levels of heavy metals in Industrial Effluents. Heavy metal levels in the sludge are now below the maximum required levels stipulated by the DNH&PD guidelines (see Appendix 13). Accordingly if the sludge were to be pasteurised it would classify as a Type D sludge and its disposal would not be restricted. Costs which are currently incurred in transporting the un-pasteurised sludge to the Refuse Disposal Site could be beneficially employed to provide pasteurisation. The production of a Type D sludge would prove attractive to those in agriculture who in all likelihood would provide transportation for the sludge.

In considering viable pasteurisation processes for Athlone, any process which relies on biogas as an energy source (such as the steam pasteurisation process employed at Cape Flats) cannot realistically be considered as a large proportion of the biogas generated at Athlone in the existing anaerobic digestion plant is required to fuel the gas engine which generates power for aeration of the activated sludge plant. Dual Digestion is however, a process which pasteurises the sludge and increases the quantity of biogas available for external use; because the heat requirements for mesophilic digestion are provided by the sensible heat of the aerobic sludge. As the dual digestion plant already exists at Athlone, and therefore the capital costs paid for, maximum benefit should be derived from its operation. The discussion below, outlines the proposed method of operation i.e. aeration of the aerobic reactor with a combination of air and pure oxygen injection to meet this objective.

9a.2 Combining the Benefits of Both the Air and Pure Oxygen Aerobic Processes

Whilst operating costs for the Dual Digestion process using air are reasonable, and a significant improvement in performance is achieved during periods of foaming, the primary disadvantage of the process is that due to the cooling effects of the nitrogen, relatively long retention times are required in the aerobic reactor, as a consequence; (1) significant VS destruction takes place in the aerobic reactor which adversely affects biogas production in the anaerobic digester, and (2) insufficient sensible heat is transferred to the digester to maintain mesophilic temperatures.

Increasing the heat input to the aerobic reactor would allow shorter retention times (i.e. increased treatment capacity) whilst maintaining operation at thermophilic temperatures. Additional heat can be generated biologically by injecting pure oxygen. Because of the low vent gas flow rates associated with oxygen injection, there is a negligible increase in the heat losses from the system.

In the Dual Digestion process using pure oxygen alone for oxygenation, because the heat losses with the vent gas are negligible, retention times as short as 1 day can be achieved whilst maintaining the aerobic reactor operating temperature at 60°C, and all the heat requirements for mesophilic anaerobic digestion are provided whilst increasing the quantity of biogas available for external use. The disadvantage of the process is the high operating costs involved with using pure oxygen. By combining the two processes, i.e. by oxygenating the aerobic reactor with air whilst supplementing with pure oxygen injection, the advantages of both processes can be combined and the disadvantages of both processes greatly reduced. In addition to increasing the treatment capacity of the air process and reducing the pure oxygen supply rate, the other main benefits are:

- Maintenance of operating temperatures in the thermophilic temperature range with a positive foam layer resulting in a consistently high transfer efficiency of oxygen from the air stream.

- Greater control over the reactor temperature because reactor temperature is known to respond rapidly to any change to the pure oxygen supply rate (Messenger *et al*, 1992).

The importance of the first point should not be overlooked. Foaming is fundamental to achieving enhanced performance of the aerobic reactor using air. By employing pure oxygen injection to raise the sludge temperature to levels (i.e. in excess of 60°C) which the air alone process was unable to achieve, due to excessive foaming, significantly greater benefits could be derived from the foam with regard to the performance of the aerobic reactor. In order to gain a better understanding of the reasons for foam formation, the report compiled after the completion of phase I (Pitt and Ekama, 1993) recommended and outlined a procedure for a pilot plant scale study into the foaming phenomenon. The results of this investigation are contained in a report by Samson (1995).

9a.3 Predicted Required Oxygen Supplementation Rates

The rates of oxygen supplementation required to reduce the minimum required retention time predicted for the air alone process were calculated in Section 5.4. Rates were calculated on the assumption that the oxygen transfer efficiency is 80% (a conservative estimate). Accepting that the reactor is not foaming and that the air flow rate is set at 760 m^3, Table 9a.1 summarises the required oxygen supplementation rates for winter and summer operation of the reactor at a temperature of 50° and 60°C.

Table 9a1 Estimated Pure Oxygen Supplementation Rates ($kg(O_2)/h$) for Summer and Winter Operation

Season	$T(SL)_r$ = 50°C		$T(SL)_r$ = 60°C	
	R_h = 1.5d	R_h = 1.0d	R_h = 1.5d	R_h = 1.0d
Summer (25°C)	20	46	50	85
Winter (15°C)	43	78	73	118

It is important to note that the pure oxygen supplementation rates given above in Table 9a.1 should be used only as guidelines. With the aerobic reactor foaming, the supplementation rates would be significantly reduced and therefore need only to be

considered as absolute maximum required rates (see Table 5.5 in Chapter 5 for the predicted rates for foaming conditions).

Supplementing the aeration of the aerobic reactor with pure oxygen will significantly increase the oxygen transfer rate ***OTR***, which under oxygen limiting conditions will be equal to the oxygen utilisation rate ***OUR***. However, there is a limit to the extent to which the ***OUR*** can be increased. At some point, there will be insufficient substrate available for utilisation by the bacteria to match the available oxygen, with the result that the reactor will become substrate limited rather than oxygen limited. At Milnerton, the maximum oxygen utilisation rate which could be achieved before the ***OUR*** became limited was 0.40 $kg(O_2)/m^3.h$ at a feed solids concentration of 40 kg/m^3 (Messenger *et al*, 1992). If a similar limitation to the ***OUR*** can be expected at Athlone (due to the similarity in sludge types), then the limit on the maximum oxygen supplementation rate would be approximately 58 $kg(O_2)/h$. It should be noted that if the OUR_{max} at Milnerton was limited as a result of insufficient substrate being available then OUR_{max} could be increased by increasing the feed sludge concentration.

Consideration must also be given to the fact that with the increased oxygen transfer rate ***OTR*** employed to obtain a reduction in the aerobic retention time, the retention time may be reduced to such an extent that the activity of the biomass (i.e. quantity of thermophilic bacteria) is reduced, which will then cause the reactor to become biomass limited.

Finally, there will be a limitation to the maximum oxygen transfer rate OTR_{max} which can be effected by the pure oxygen injection device. This limitation is principally dependent on the solubility of the pure oxygen at the point of injection. Consideration must therefore be given to the physical conditions (temperature, pressure and sludge flow rate) at this point. The maximum oxygen supplementation rate predicted for injection at a pressure of 2280mmHg with the sludge at 60°C is $60kg(O_2)/h$ (see Appendix 15).

9a.4 Practical Aspects of Pure Oxygen Supplementation

The aerobic reactor is currently aerated with air, supplied by a liquid ring compressor, which is passed into the reactor through a set of course bubble diffusers set in the base of the reactor. It is envisaged that pure oxygen for supplementation would be injected into the sludge (with Venturi type injector) at a point immediately downstream of the mixing pumps on the sludge recirculation line. This particular point is chosen for the following reasons:

- Oxygen injection and subsequent utilisation must take place away from the main air flow through the reactor; any contact with the air flow would reduce the transfer efficiency of the pure oxygen.

- Oxygen injection must be carried out after passage through the mixing pump; injection upstream of the pump may result in cavitation in the pump causing damage

Liquid oxygen would be stored on site in a cryogenic storage vessel, the size of which would be dependent upon required supplementation rates (see Table 9a.1) and the frequency of delivery. A suitable power point and supporting concrete base would be required for the storage vessel. It is appropriate to note at this stage the recent developments which have taken place regarding the technology of on site oxygen generation (Vacuum Swing Absorption, VSA) and its availability in South Africa (see Appendix 14). The VSA plant is ideally suited to the dual digestion process for the following reasons:

- The dual digestion process does not require high purity oxygen, as such oxygen concentrations of the order of 90% would be acceptable which would allow the VSA plant to provide a relatively high cost effective output of oxygen.

- Maximum benefit is derived from the VSA plant when under continuous operation which would be the case for the aerobic reactor.

The technology is available in South Africa and the leading gas supply companies are in a position to manufacture and supply VSA plants locally to user specifications. The feasibility of employing VSA should be investigated. If an appropriately sized VSA plant is not available for testing during the trial period, then the results obtained with pure liquid oxygen injection could be used to determine the size of VSA plant required in future, if VSA were to prove viable.

In addition to the required oxygenation equipment, the following work needs to be carried out to the existing plant before the investigation can commence:

- Appropriately sized motors need to be fitted to the mixing pumps on the sludge recirculation line. Two 22kW motors are available from the de-commissioned Cape Flats Reclamation Plant.

- All the current instrumentation needs to be checked and recalibrated

- The inside of the aerobic reactor must be clean. Domes should be intact etc.

- The sludge in the anaerobic digester should be tested to make sure that the sludge is still active. If need be it should be fed in the period leading up to the investigation.

- The feasibility of pre-heating the feed sludge through the existing heat exchange system should be investigated.

This last point is of particular importance, as the treatment capacity of the aerobic reactor can also be increased by pre-heating the feed sludge. Pre-heating in conjunction with providing pure oxygen supplementation, would significantly reduce the quantities of pure oxygen required to reach a specific temperature (see Section 8.3). With biogas providing the fuel requirements for pre-heating, it is necessary therefore to consider a future sludge treatment programme which makes efficient use of the biogas generated in the anaerobic digesters. Consideration must be given to biogas requirements for mixing in the anaerobic digester(s) and as fuel for the gas engine. Any excess biogas could then be used for pre-heating the feed sludge.

To satisfy the monitoring requirements of the investigation, the following items of monitoring equipment should be obtained:

- A portable oxygen analyzer to determine the oxygen concentration in the vent gas, and consequently the oxygen utilisation rate, or preferably an in line oxygen analyzer for continuous vent gas monitoring.

- A DO meter, capable of operation at thermophilic temperatures, would be required in the aerobic reactor; to indicate the onset of OUR limitation.

- Gas meters on both the inlet and the outlet to the anaerobic digester. This will enable the heat losses from the digester to be assessed and to estimate the biogas production.

9a.5 Aims and Objectives of the Investigation

The principal aim of the investigation would be to assess the viability of the dual digestion process using air with pure oxygen supplementation to satisfactorily treat a high proportion (if not all) of the gravity thickened sludge produced at Athlone and to determine the annual operating costs of such treatment. Specific objectives of the investigation would be to:

- Determine future operating costs and compare with other pasteurisation processes.
- Determine the transfer efficiency of the oxygen injection equipment.
- Examine the effects of foam during oxygen injection
- Establish whether the heat requirements for mesophilic digestion are met.
- Determine the maximum possible treatment capacity
- Determine the effect of aerobic treatment on subsequent biogas production
- Examine the practicability of the process
- Determine the minimum feed sludge concentrations to prevent substrate limitation
- Develop a model to describe the kinetics of growth and VS removal
- Investigate the options for efficiently utilising the biogas produced.

To achieve the above objectives, it is envisaged that the investigation would last for a period of 9-12 months, including a summer and winter season. The reactor would be operated at a number of different steady state periods, each differing from each other with respect to retention time and reactor temperature.

9a.6 Theoretical Considerations

In addition to meeting the aims and the objectives of the investigation detailed in Section 6a.5 above, the investigation also provides the opportunity to examine a specific aspect of aerobic bio-kinetics, that is VS destruction at short retention times.

At retention times in excess of 3 days, the predominant bacterial activity which takes place within the aerobic reactor is that of VS destruction. The rate of volatile solids destruction can accordingly, be estimated from an application of VS destruction kinetics (Andrews and Kambhu, 1971). These kinetics can then be used to predict the minimum required feed sludge concentration to ensure substrate limitation does not take place for a specified retention time; values calculated from the model agree closely with those observed during the evaluation.

At the short retention times encountered when using pure oxygen the predominant biological activity is no longer VS destruction but also organism growth (Messenger *et al*, 1992). As such, the kinetics of Andrews and Kambhu are no longer valid for predicting VS removal and conditions for substrate limitation. To date, no work/papers have been published on modelling VS destruction at short retention times, as such, it is intended to use the data derived from the trial period to develop a satisfactory model, which will consider the growth kinetics of thermophilic organisms, or at least be able to make a satisfactory prediction of VS removal.

9a.7 Conclusions

The existing dual digestion plant at Athlone has the potential to pasteurise a substantial portion of the primary sludge produced at Athlone simply by supplementing the aeration of the aerobic reactor with pure oxygen. It is therefore recommended that the investigation into pure oxygen supplementation as described above be carried out on the Athlone dual digester as part of Athlone's sludge treatment research, development, and improvement programme (labled phase II of the investigation into the dual digestion process).

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Samson KA (1995) Foam formation in the air oxygenated thermophilic aerobic reactor of the dual digestion process: A pilot scale study. SSB File Ref N° CB.2/S13, Scientific Services Branch, City Engineer's Department, Cape Town City Council, PO Box 1694, Cape Town, South Africa.

APPENDIX 10

DETERMINATION OF SLUDGE FLOW RATE IN THE RECIRCULATION LINE: PHASE II

10a.1 Introduction

The modifications which were effected to the sludge recirculation line as part of the requirements for efficient pure oxygen transfer are illustrated in Figure 10a.1 below. In order to determine the flow rate of sludge through the line and the velocities through the Venturi and at the point of discharge, all of which are important parameters which affect the transfer efficiency of the injection device, it is necessary to calculate the pressure drop across the length of the line at different flow rates (velocities). This will produce a "system head loss curve", which superimposed on the pump curve(s) will enable the flow rate and consequently the velocities to be determined.

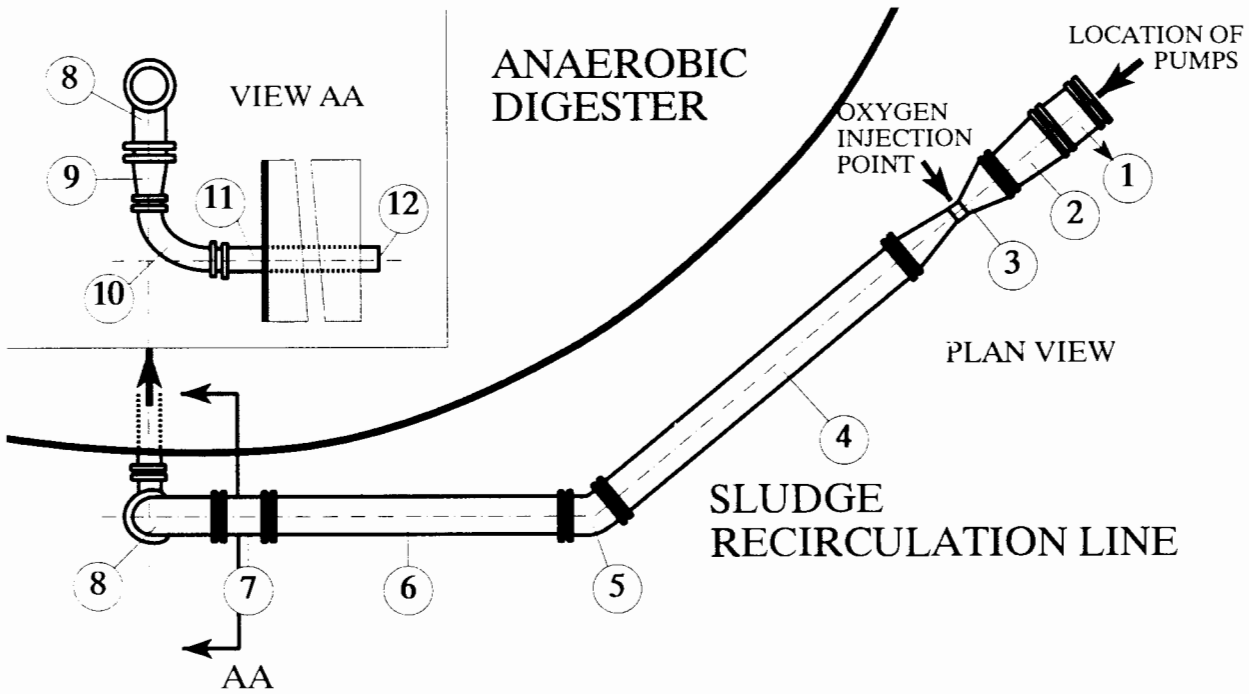


Figure 10a.1 Schematic of the Sludge Recirculation Line as Constructed for Phase II of the Evaluation Period (The Pipe Fittings are Numbered for Later Reference, See Section 10a.3 below).

10a.2 The Pumping of Liquids

Bernoulli's Theorem: Bernoulli's theorem is an expression of the conservation of energy law, applied to the flow of a fluid in a system. Bernoulli's equation for the total energy at any particular point in the system, above an arbitrary base line, is as follows:

$$H = z + \frac{P}{\rho g} + \frac{v^2}{2g} \quad \text{...m} \quad (10a.1)$$

where:

H = total head (m)
z = static head (m)
P = gas pressure above liquid (kg/ms^2)
$P/\rho g$ = pressure head (m)
v = liquid velocity (m/s)
$v^2/2g$ = velocity head (m)

If friction losses are neglected and no energy is added to, or taken from, the system, then the total head H will be constant at any point in the system. Usually, however, a pump is provided to supply energy, or head, to the flowing liquid in order to overcome the head losses due to friction. If a pump is placed between points 1 and 2 in a pipeline, the energy balance is given by:

$$z_1 + \frac{P_1}{\rho g} + \frac{v_1^2}{2g} = z_2 + \frac{P_2}{\rho g} + \frac{v_2^2}{2g} + \Delta h - h_f \quad \text{...m} \quad (10a.2)$$

where:

Δh = head imparted by the pump (m)
h_f = head loss due to friction (m)

System Heads: For the pumping system, associated with the Athlone aerobic reactor (referred to as the sludge recirculation line), the important system heads to consider are the suction, discharge and total heads (see Figure 10a.2). The centreline of the pump is chosen as the base line. The suction head is given by:

$$h_s = z_s + \frac{P_s}{\rho g} - h_{fs} \quad \text{...m} \quad (10a.3)$$

and the discharge head is given by:

$$h_d = z_d + \frac{P_d}{\rho g} - h_{fd} \quad \text{...m} \quad (10a.4)$$

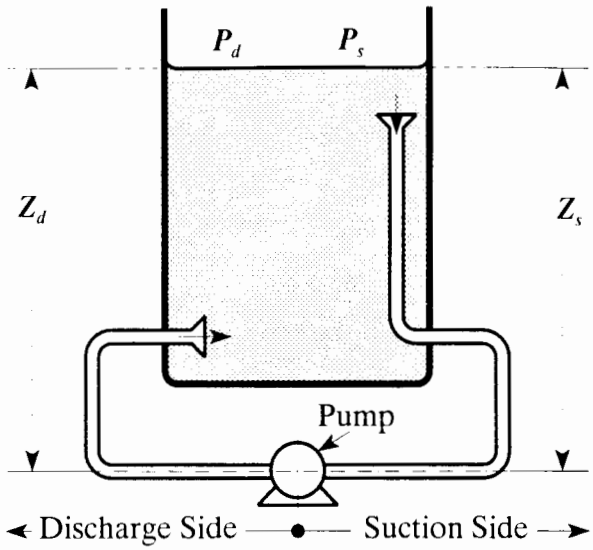


Figure 10a.2 Schematic of the Aerobic Reactor Pumping System Showing the Static Heads z_s and z_d and the Pressure(s) on the liquid P_s and P_d.

The total head Δh which the pump is required to impart to the flowing sludge liquid is the difference between the discharge and suction heads viz:

$$\Delta h = h_d - h_s \quad \text{...m} \quad (10a.5)$$

Substitution of Eq's 10a.3 and 10a.4 into the above yields:

$$\Delta h = (z_d - z_s) + \frac{(P_d - P_s)}{\rho g} + (h_{fd} + h_{fs}) \quad \text{...m} \quad (10a.6)$$

For the pumping system at Athlone, the static head and the pressure head are the same on both the delivery and suction sides (see Figure 10a.2 above) i.e. $z_s = z_d$ and $P_s = P_d$. Consequently, the total head Δh imparted by the pump is equal to the sum of the suction and delivery friction head losses $(h_{fd} + h_{fs})$ viz:

$$\Delta h = (h_{fd} + h_{fs}) \quad \text{...m} \quad (10a.7)$$

Friction Head Losses: From D'Arcy's formula, the friction head loss h_f through a straight pipe of length L and diameter D is given by:

$$h_f = f.\frac{L}{D}.\frac{v^2}{2.g} \quad \text{...m} \quad (10a.8)$$

where:

f = friction factor (-)

The volumetric flow rate of sludge Q and the mean velocity of the flow v are linked by the following equation:

$$v = \frac{Q}{3600} \cdot \frac{4}{\pi . D^2} \quad \text{...m/s} \quad \text{(10a.9)}$$

where:

Q = volumetric flow rate of sludge (m^3/h)

3600 = conversion factor (s/h)

D'Arcy's formula (Eq 10a.8) can therefore be rewritten as:

$$h_f = 6.3 \times 10^{-8} . \frac{f . L . Q^2}{\pi^2 . D^5} \quad \text{...m} \quad \text{(10a.10)}$$

The friction factor f is a function of two dimensionless terms: The Reynolds number Re, and the ratio e/D, where e is a length representing the surface roughness and D is the diameter of the pipe. The Reynolds number Re is given by the following equation:

$$Re = \frac{v . D . \rho}{\mu} \quad \text{....} \quad \text{(10a.11)}$$

where:

μ = absolute dynamic viscosity of the sludge (kg/m.s)

Both the sludge mass density ρ and the viscosity of the sludge μ are taken to be that of water. At the typical operating temperature of 60°C (taken from Mayhew and Rogers, 1977) viz:

ρ = 983 kg/m^3

μ = 0.467 x 10^{-3} kg/m.s

For a sludge flow rate Q in the recirculation line ranging between 200 and 1000 m^3/h, and with the pipe diameters D ranging between 0.136 and 0.30 m, the minimum and maximum expected velocities v in the pipeline are 1.1 and 19.1 m/s respectively. Under such flow conditions, the Reynolds number falls within the range (from Eq 10a.11):

$$6 \times 10^5 < Re < 5 \times 10^6 \quad \text{...} \quad \text{(10a.12)}$$

At such high Reynolds numbers, the flow regime is considered to be turbulent (Coulsen and Richardson, 1977), and the friction factor f is independent of Re (which is why it is not necessary to take into account the increased viscosity and density due to the presence of sludge mass in the water) and depends only on the relative roughness of the

pipewalls e/D. The absolute roughness e of the pipework is taken to be 0.00026 m (that quoted for cast iron). The relative roughness e/D then falls within the range:

$$0.001 < \frac{e}{D} < 0.002 \quad \text{... (10a.13)}$$

From the pipe friction factor chart f versus Re (Coulsen and Richardson, 1977), for the range of relative roughness e/D quoted above, a friction factor of $f = 0.015$ is obtained. Therefore, from D'Arcy's formula (Eq 10a.10), the friction head loss h_f through a straight pipe of length L and diameter D at a sludge flow rate of Q is given by:

$$h_f = 9.4 \times 10^{-10} . \frac{L . Q^2}{\pi^2 . D^5} \quad \text{...m (10a.14)}$$

To determine the friction head loss h_f through either a valve or specific fitting, the expression for h_f is written in terms of the resistance coefficient K. Standard formulae are employed to calculate K based on the geometry of the particular fitting (Crane 1965).The friction head loss is given by:

$$h_f = K . \frac{v^2}{2.g} \quad \text{...m (10a.15)}$$

where:

K = resistance coefficient

In terms of the volumetric flow rate Q Eq 10a.15, can be rewritten as follows (after substitution of Eq 10a.9):

$$h_f = 6.3 \times 10^{-8} . \frac{K . Q^2}{\pi^2 . D^4} \quad \text{...m (10a.16)}$$

10a.3 Calculation of the Head Loss Due to Friction in the Recirculation Line

The total head loss due to friction h_f^{Total} in the sludge recirculation line is determined by summing the friction head losses through the pipework on the discharge side h_{fd}^{Total} and the suction side h_{fs}^{Total} viz:

$$h_f^{Total} = h_{fd}^{Total} + h_{fs}^{Total} \quad \text{...m (10a.17)}$$

The total head loss due to friction on the discharge side h_{fd}^{Total} is determined by summing the friction head losses through each of the individual pipe fittings on the discharge side (see Figure 10a.1) viz:

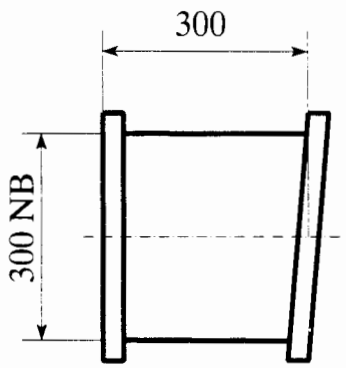
$$h_{fd}^{Total} = \sum_{i=1}^{12} h_{fd}^{i} \quad \text{...m} \quad (10a.18)$$

where:

h_{fd}^{i} = the head loss through the i^{th} fitting (m)

The calculation for each fitting is as follows:

Fitting No. 1 Straight Pipe ϕ300mm × 0.3m:

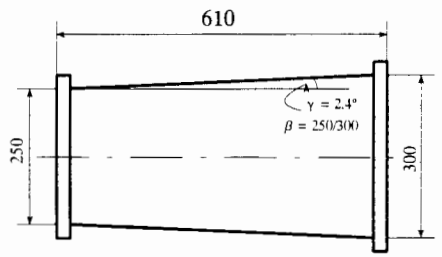


D = 0.300m L = 0.300m

$$h_{fd}^{1} = 9.4 \times 10^{-10} . \frac{0.300 \times Q^2}{\pi^2 \times 0.300^5}$$

$$h_{fd}^{1} = 1.18 \times 10^{-8} . Q^2$$

Fitting No. 2 Reducer 300/200mm:

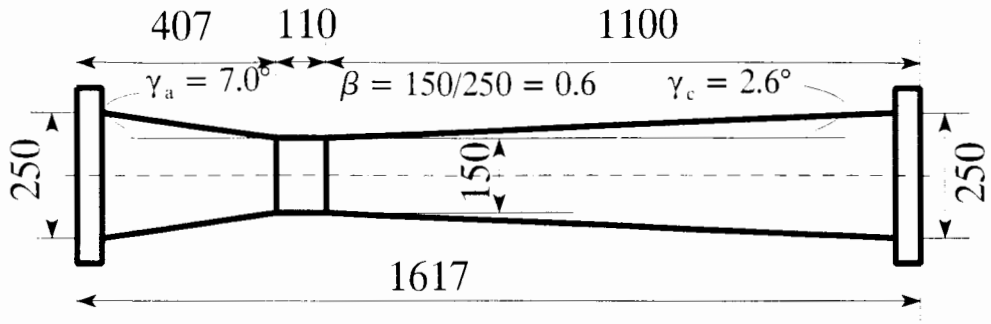


D = 0.300m $K = \frac{0.8 \sin(\gamma).(1-\beta^2)}{\beta^4} = 0.021$

$$h_{fd}^{2} = 6.3 \times 10^{-8} . \frac{0.021 \times Q^2}{\pi^2 \times 0.300^4}$$

$$h_{fd}^{2} = 1.65 \times 10^{-8} . Q^2$$

Fitting No. 3 Venturi:



Converging Section:

$D = 0.250\text{m}$ $\quad K = \dfrac{0.8\ \sin(\gamma_a).(1-\beta^2)}{\beta^4} = 0.481$

$$h_{fd}^{3a} = 6.3\times10^{-8}.\frac{0.481\times Q^2}{\pi^2\times0.250^4} = 7.86\times10^{-7}.Q^2$$

Straight Section:

$D = 0.150\text{m}$ $\quad L = 0.110\text{m}$

$$h_{fd}^{3b} = 9.4\times10^{-10}.\frac{0.110\times Q^2}{\pi^2\times0.150^5} = 1.38\times10^{-7}.Q^2$$

Diverging Section:

$D = 0.150\text{m}$ $\quad K = \dfrac{2.6\ \sin(\gamma_c).(1-\beta^2)}{\beta^4} = 0.37$

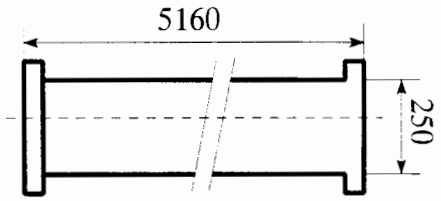
$$h_{fd}^{3c} = 6.3\times10^{-8}.\frac{0.37\times Q^2}{\pi^2\times0.150^4} = 4.66\times10^{-6}.Q^2$$

Total Pressure Drop Across Venturi

$$h_{fd}^{3} = h_{fd}^{3a} + h_{fd}^{3b} + h_{fd}^{3c} = 5.58\times10^{-6}.Q^2$$

Fitting No. 4 Straight Pipe ϕ250mm × 5.16m:

$D = 0.250\text{m}$ $\quad L = 5.16\text{m}$



$$h_{fd}^{4} = 9.4\times10^{-10}.\frac{5.16\times Q^2}{\pi^2\times0.250^5}$$

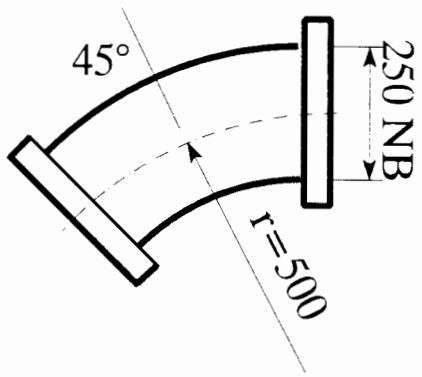
$$h_{fd}^{4} = 5.03\times10^{-7}.Q^2$$

Fitting No. 5 **45° Mitre Bend ϕ250mm:**

D = 0.250m $\quad K = 15 \times f = 0.23$

$$h_{fd}^5 = 6.3 \times 10^{-8} . \frac{0.23 \times Q^2}{\pi^2 \times 0.250^4}$$

$$h_{fd}^5 = 3.76 \times 10^{-7} . Q^2$$

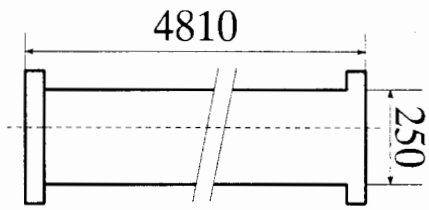


Fitting No. 6 **Straight Pipe ϕ250mm × 4.81m:**

D = 0.250m $\quad L$ = 4.81m

$$h_{fd}^6 = 9.4 \times 10^{-10} . \frac{4.81 \times Q^2}{\pi^2 \times 0.250^5}$$

$$h_{fd}^6 = 4.69 \times 10^{-7} . Q^2$$

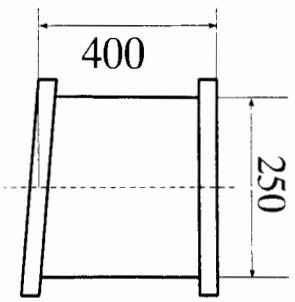


Fitting No. 7 **Straight Pipe ϕ250mm × 0.4m:**

D = 0.250m $\quad L$ = 0.400m

$$h_{fd}^7 = 9.4 \times 10^{-10} . \frac{0.400 \times Q^2}{\pi^2 \times 0.250^5}$$

$$h_{fd}^7 = 3.90 \times 10^{-8} . Q^2$$

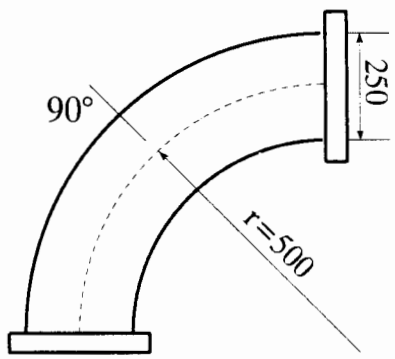


Fitting No. 8 **90° Bend ϕ250mm:**

D = 0.250m $\quad K = 15 \times f = 0.23$

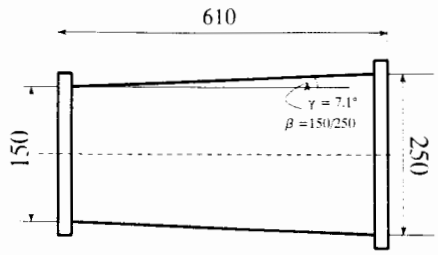
$$h_{fd}^8 = 6.3 \times 10^{-8} . \frac{0.23 \times Q^2}{\pi^2 \times 0.250^4}$$

$$h_{fd}^8 = 3.76 \times 10^{-7} . Q^2$$



Fitting No. 9 **Reducer 250/150mm:**

$D = 0.250\text{m}$ $\quad K = \dfrac{0.8\ \sin(\gamma).(1-\beta^2)}{\beta^4} = 0.490$

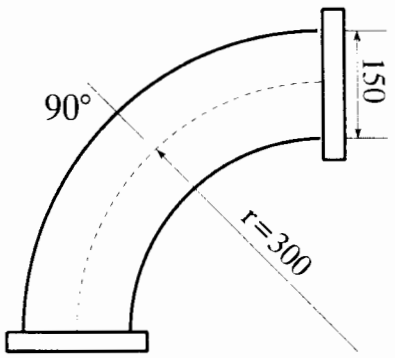


$$h_{fd}^{9} = 6.3\times10^{-8}.\frac{0.490\times Q^2}{\pi^2\times0.250^4}$$

$$h_{fd}^{9} = 8.00\times10^{-7}.Q^2$$

Fitting No. 10 **90° Bend ϕ150mm:**

$D = 0.150\text{m}$ $\quad K = 12\times f = 0.18$

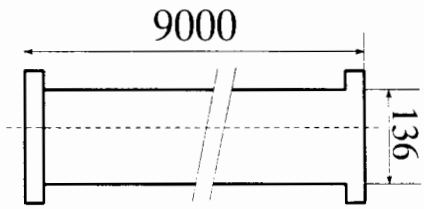


$$h_{fd}^{10} = 6.3\times10^{-8}.\frac{0.18\times Q^2}{\pi^2\times0.150^4}$$

$$h_{fd}^{10} = 2.27\times10^{-6}.Q^2$$

Fitting No. 11 **Straight Pipe ϕ136mm $\times$ 9.0m:**

$D = 0.136\text{m}$ $\quad L = 9.0\text{m}$

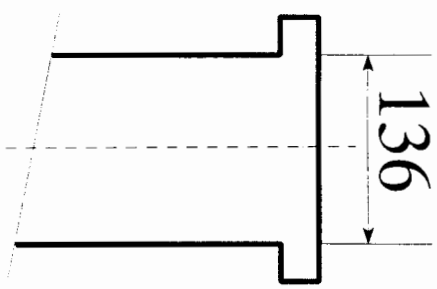


$$h_{fd}^{11} = 9.4\times10^{-10}.\frac{9.0\times Q^2}{\pi^2\times0.136^5}$$

$$h_{fd}^{11} = 1.84\times10^{-5}.Q^2$$

Fitting No. 12 **Pipe Exit ϕ136mm:**

$D = 0.136\text{m}$ $\quad K = 1.0$



$$h_{fd}^{12} = 6.3\times10^{-8}.\frac{1.0\times Q^2}{\pi^2\times0.136^4}$$

$$h_{fd}^{12} = 1.87\times10^{-5}.Q^2$$

The total head loss due to friction on the discharge side h_{fd}^{Total}, obtained by summing the individual friction losses (Eq 10a.18), is as follows:

$$h_{fd}^{Total} = \sum_{i=1}^{12} h_{fd}^{i} = 4.75 \times 10^{-5}.Q^2 \quad \text{...m} \quad \text{(10a.19)}$$

As the total head loss due to friction on the suction side h_{fs}^{Total} is insignificant in comparison to that calculated for the discharge side, a simple estimate of the head loss is sufficient. It is assumed that the pipework on the suction side consists of an equivalent 20m length of straight pipe of diameter ϕ300mm

<u>Suction Side</u> Equivalent Straight Pipe ϕ300mm × 20.0m:

D = 0.300m L = 20.0m

$$h_{fs}^{Total} = 9.4 \times 10^{-10}.\frac{20.0 \times Q^2}{\pi^2 \times 0.300^5}$$

$$h_{fs}^{Total} = 7.84 \times 10^{-7}.Q^2 \quad \text{...m} \quad \text{(10a.20)}$$

The overall friction head loss through the sludge recirculation line h_f^{Total} is obtained by summing the head loss through the discharge side h_{fd}^{Total} and the suction side h_{fs}^{Total} :

$$h_f^{Total} = 4.83 \times 10^{-5}.Q^2$$

Equation 10a.21 System Head Loss Curve: The Total Friction Head Loss h_f^{Total} Through the Sludge Recirculation Line (m).

10a.4 Compilation of the Operating Pump Curves

The same two pumps were used throughout both evaluation periods (phase I and phase II). However, the impellers, motors, and configurations were changed on occasion.

Phase I: During phase I (when operation was with air alone), the pumps were positioned in parallel. Each pump was operated alternately at fixed time intervals (usually 4 hrs). The pumps were fitted with 7.5kW motors and 408mm diameter impellers. Because of the negligible friction losses in the recirculation pipeline during phase I, the rate of sludge pumping in the recirculation line (estimated from the pump curve) was in excess of $1000m^3/h$.

Phase II: For the proposed incorporation of pure oxygen injection during phase II, the sludge recirculation line was modified (see Figure 10a.1). The motors and impellers on the pumps were changed to try and obtain the correct pumping conditions for successful pure oxygen injection; it was considered that a sludge flow rate of $700m^3/h$, at a system head of 20m, were required. Subsequently, two further modifications were made to the pumping set-up during phase II to try and achieve the correct pumping conditions. The three sets of pumping conditions which existed during phase II are described in Table 10a.1 below.

Table 10a.1 Pumping Conditions During Phase II

Operating Period (Day No's)	No. of Pumps	Operational Configuration	Pump 1 (55kW)		Pump 2 (75kW)	
			Frequency (rpm)	Impeller Size (mm)	Frequency (rpm)	Impeller Size (mm)
1-25	1	-	1470	280	-	-
26-40	2	Parallel	1470	280	1470	280
74-152	2	Series	1470	330	1617	350

The oxygen transfer efficiency which the injection device is capable of effecting is directly influenced by the flow characteristics in the sludge recirculation line. Consequently, it is necessary to know the flow conditions in the recirculation line (in terms of flow rate and pressure) for each of the three pumping conditions described in Table 10a.1 above. This is achieved by superimposing the relevant pump characteristic curves onto the system characteristic curve. The situation is made more complex by the fact that for the last two pumping periods, two pumps were in operation, firstly in parallel and then in series.

Manufacturer's Pump Curve: The pump curve supplied by the Manufacturer of the pump(s) used in the sludge recirculation line is displayed in Figure 10a.3 below.

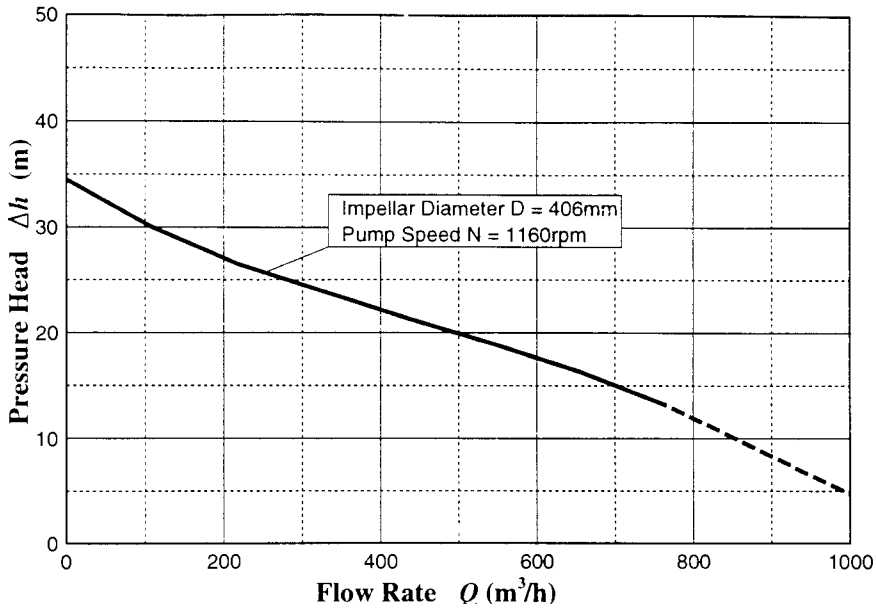


Figure 10a.3 Manufacturer's Supplied Pump Characteristic Curve for the Pump(s) used in the Sludge Recirculation Line

Pumping Condition 1: (days 1-25)

At the start of phase II, between days 1 and 25, one pump was in operation. A 55kW motor was fitted to the pump, operating at 1470rpm. The size of the impeller measured 280mm in diameter. Figure 10a.4 below shows the appropriate pump curve with system curve superimposed.

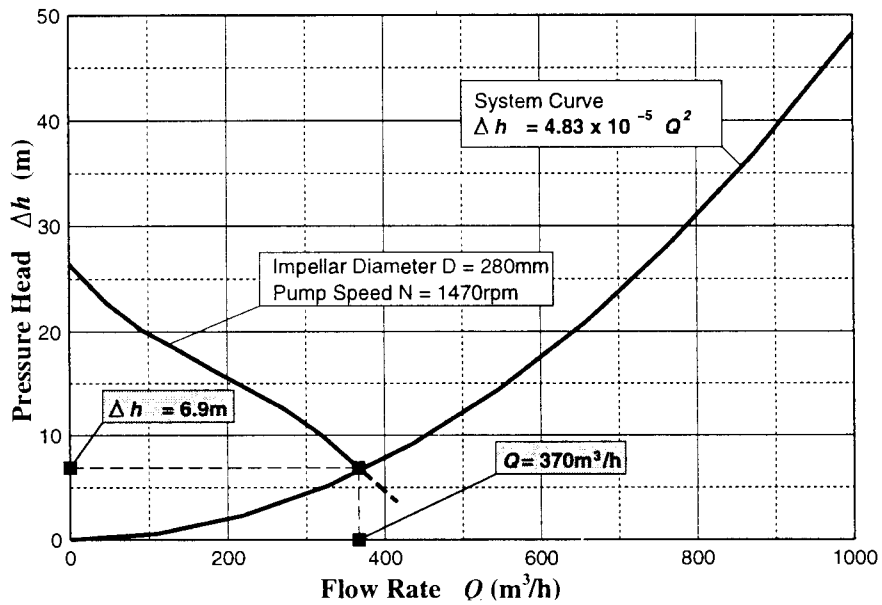


Figure 10a.4 Pump and System Curves for the Initial Set of Pumping Conditions During Phase II: 1 Pump in Operation N= 1470rpm D=280mm

The above pump curve, and those displayed hereafter, were derived from the original pump curve by applying the following formulae, which relate the head loss Δh and flow rate Q between two different sets of conditions (denoted by subscripts 1 and 2):

$$\Delta h_2 = \Delta h_1 \cdot \left(\frac{N_2}{N_1}\right)^2 \cdot \left(\frac{D_2}{D_1}\right)^2 \quad \text{...m} \quad (10a.22)$$

and

$$Q_2 = Q_1 \cdot \left(\frac{N_2}{N_1}\right) \cdot \left(\frac{D_2}{D_1}\right)^3 \quad \text{...m}^3\text{/h} \quad (10a.23)$$

where:

N = the speed of the pump (rpm)

D = the diameter of the pump impeller (mm)

Pumping Condition 2: (days 26-40)

During phase II, between days 26 and 40, two pumps were operated in parallel. Both pumps operated at 1470rpm, and the both had the same size impeller i.e. 280mm. One pump was fitted with a 55kW motor, the other fitted with a 75kW motor. Both pumps have the same pump characteristic curve, as illustrated in Figure 10a.4 above. The head and flow rate developed during parallel operation is determined graphically in Figure 10a.5 below:

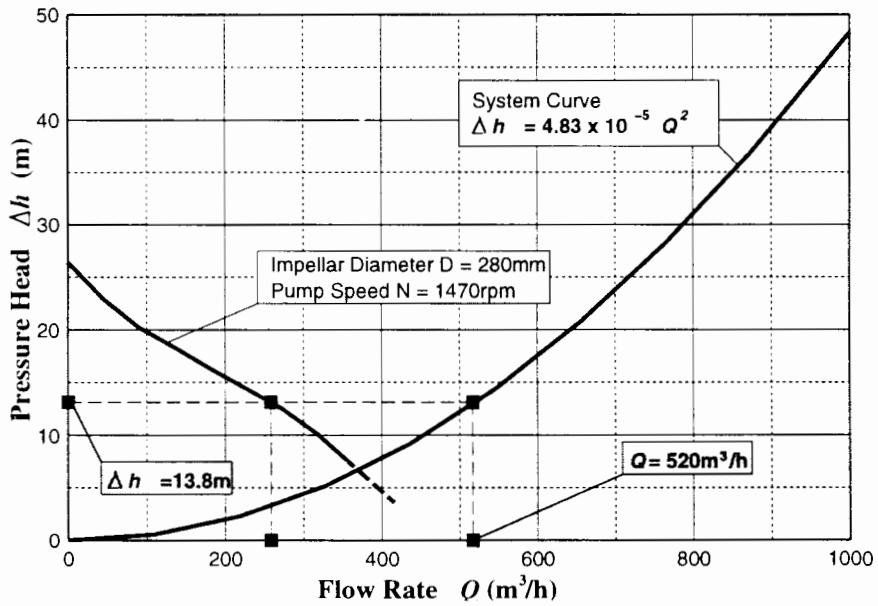


Figure 10a.5 Pump and System Curves for the Second Set of Pumping Conditions During Phase II: 2 Pumps in Parallel Operation, For both Pumps N=1470rpm and D=280mm

In parallel operation, the head developed by each pump must be the same as the system head. Referring to Figure 10a.5, a horizontal line (representing constant head) is drawn on the graph which intersects the pump and system curves, and gives a flow rate Q_T equal to twice Q_a.[1]

Pumping Condition 3: (days 74-152)

During phase II, between days 74 and 152, the two pumps were operated in series. The first pump was operated at 1470rpm with an impeller size of 330mm, fitted with a 55kW motor. The second pump was operated at 1617rpm with an impeller size of 350mm, fitted with a 75kW motor. The pump characteristic curves for each pump are given in Figure 10a.6 below. The head and flow rate developed during series operation is determined graphically.

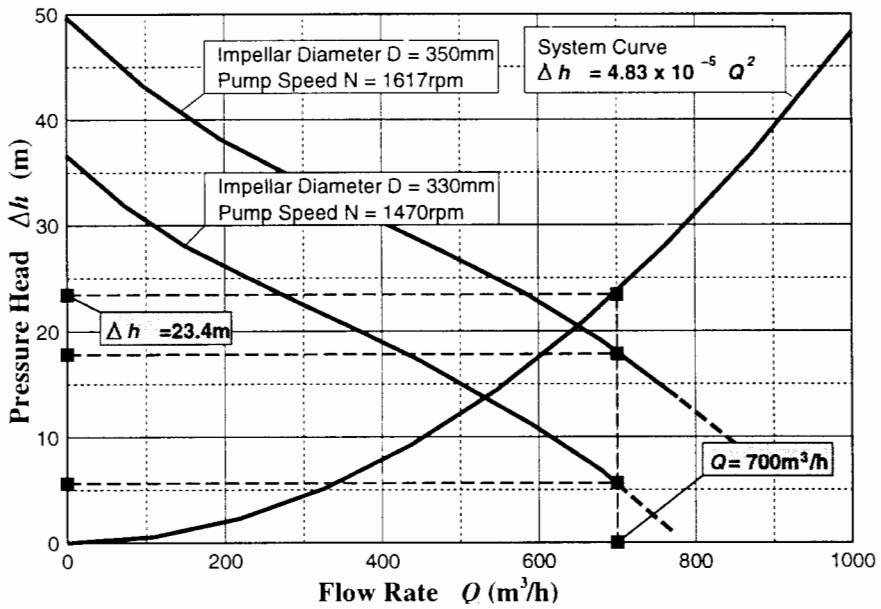


Figure 10a.6 Pump and System Curves for the Third Set of Pumping Conditions During Phase II: 2 Pumps in Series Operation, For the First Pump N=1470rpm and D=330mm. For the Second Pump N=1617rpm and D=350mm

In series operation, the heads can differ but the pump capacities (flow rate) must all be equal. Referring to Figure 10a.6, a vertical line (representing constant flow) is drawn on the graph which intersects the pump and system curves, and gives a system head Δh equal to the sum of the pump heads $\Delta h_a + \Delta h_b$.

---

[1] If two different pumps were employed in parallel (pump a and pump b), then the total flow rate Q_T would be equal to $Q_a + Q_b$.

10a.5 Summary of the Flow Characteristics for Phase II: Sludge Recirculation Line

The flow characteristics for the sludge recirculation line during phase II, which were determined graphically above in Section 10a.4, are tabulated in Table 10a.2 below.

Table 10a.2 Flow Characteristics in the Sludge Recirculation Line During Phase II

Operating Period (Day No's)	Recirculation Sludge Flow Rate (m^3/h)	Pressure Head Generated in the Recirculation Line (m)
1-25	370	6.9
26-40	520	13.8
74-152	700	23.4

10a.6 Determination of the Flow Characteristics at Different Pump Speeds

During the third pumping condition (when two pumps operated in series), the 75kW motor attached to the second pump was fitted with a variable speed drive. On day 139, the frequency of the power supply to the pump motor was varied, yielding different pump speeds (see Table 10a.3 below). The effect of the changing flow characteristics in the recirculation line on the pure oxygen transfer efficiency was then determined; full details are given in Chapter 7, Results and Discussion: Phase II (Section 7.3.3). The pump curves for each condition are shown graphically in Figure 10a.7 below.

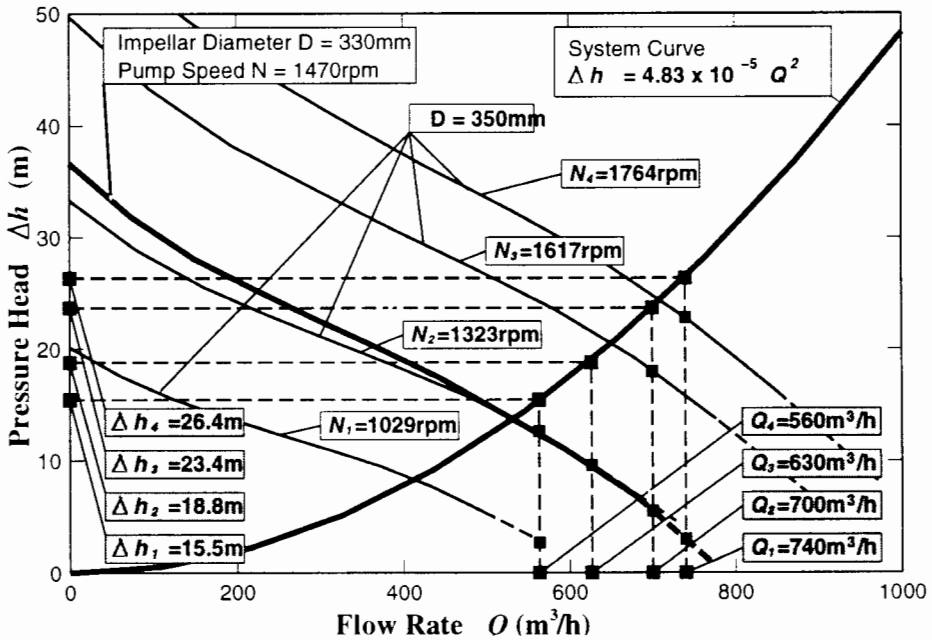


Figure 10a.7 Pump Characteristic Curves for Different Speeds of the Second Pump in Series. For the First Pump N=1470rpm and D=330mm. For the Second Pump N=1029-1764rpm and D=350mm

The flow characteristics for each pumping condition during the trial , where the speed on pump 2 was varied, are tabulated in Table 10a.3 below.

Table 10a.3 Flow Characteristics in the Recirculation Line, at Different Speeds of the Second Pump in Series. For the First Pump N=1470rpm and D=330mm. For the Second Pump N=1029-1764rpm and D=350mm

Frequency (Hz)	Pump Speed (rpm)	Flow Rate (m^3/h)	Pressure Head (m)
35	1029	560	15.5
45	1323	630	18.8
55	1617	700	23.4
60	1764	740	26.4

10a.7 Overall Summary of Pump, Flow and Oxygenation Characteristics

Table 10a.4 below contains a complete summary of all the relevent measured and estimated pump, flow and oxygenation data compiled during phase II.

Table 10a.4 Summary of the Measured and Estimated Pump, Flow, and Oxygenation Data Compiled During Phase II.

No.	Day	Flow	Velocity through Venturi	Velocity at discharge	Pressure Head Generated	Total Pressure	No of Pumps	Operation	Power in Liquid	Electrical Power Drawn	Pump Efficiency
-	-	m^3/h	m/s	m/s	m	m	-	-	kW	kW	%
1	1-25	370	5.8	7.1	6.9	23.2	1	solo	7.0	34	21
2	26-40	520	8.1	9.9	13.8	30.1	2	parallel	19.6	44	45
3	74-152	700	10.7	13.4	23.4	39.7	2	series	44.6	93	48
4	139	560	8.5	10.7	15.5	31.8	2	series	23.6	49	48
5	139	630	9.6	12.0	18.8	35.1	2	series	32.3	65	50
6	139	700	10.6	13.4	23.4	39.7	2	series	44.6	93	48
7	139	740	11.2	14.2	26.4	42.7	2	series	53.2	110	48

No.	Pump 1 (55kW Rated Motor)				Pump 2 (75kW Rated Motor)				Oxygenation Characteristics		
	Power	Speed	Impellar	Frequency	Power	Speed	Impellar	Frequency	*OSR*	*OTR*	*OTE*
-	kW	rpm	mmϕ	Hz	kW	rpm	mmϕ	Hz	$kgO/m^3.h$	$kgO/m^3.h$	$kgO/m^3.h$
1	34	1470	280	50	-	-	-	-	0.326	0.153	47.6±1.7
2	22	1470	280	50	22	1470	280	50	0.326	0.179	54.9±1.6
3	35	1470	330	50	58	1617	350	55	0.13-0.52	0.11-0.44	82.8±1.5
4	34	1470	330	50	15	1029	350	35	0.522	0.354	67.8
5	34	1470	330	50	31	1323	350	45	0.522	0.371	71.1
6	35	1470	330	50	58	1617	350	55	0.522	0.429	82.2
7	35	1470	330	50	75	1764	350	60	0.522	0.445	85.2

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Crane (1965) Flow of fluids through valves, fittings, and pipe. Technical Paper N° 410. Crane Co., 4100 S. Kedzie Ave., Chicago, Illinois, USA.

Perry RH and Chilton CH (1973) Chemical Engineer's Handbook. 5th Edition. McGraw-Hill Kogakusha.

APPENDIX 11

THE ANAEROBIC DIGESTION PROCESS: A STEADY STATE MODEL

11a.1 Introduction

A steady state model for the mesophilic and thermophilic anaerobic digestion of primary sewage sludge, pretreated in an autothermal thermophilic aerobic reactor is presented. The model will be used to predict COD (VS) destruction and methane production in the anaerobic stage of the dual digestion process.

11a.2 The Microbiology of the Anaerobic Digestion Process

The anaerobic digestion of organic matter can be considered to incorporate six distinct conversion processes performed by specific bacterial populations. The proposed reaction scheme for the anaerobic digestion of domestic sewage sludge has been described by Guger and Zehnder (1983) and is summarised in Figure 11a.1 below. The flow of substrate through the process, in the form of COD equivalents, is shown as a percentage of the initial COD. In this Figure the cell yield (growth of biomass) has been ignored.

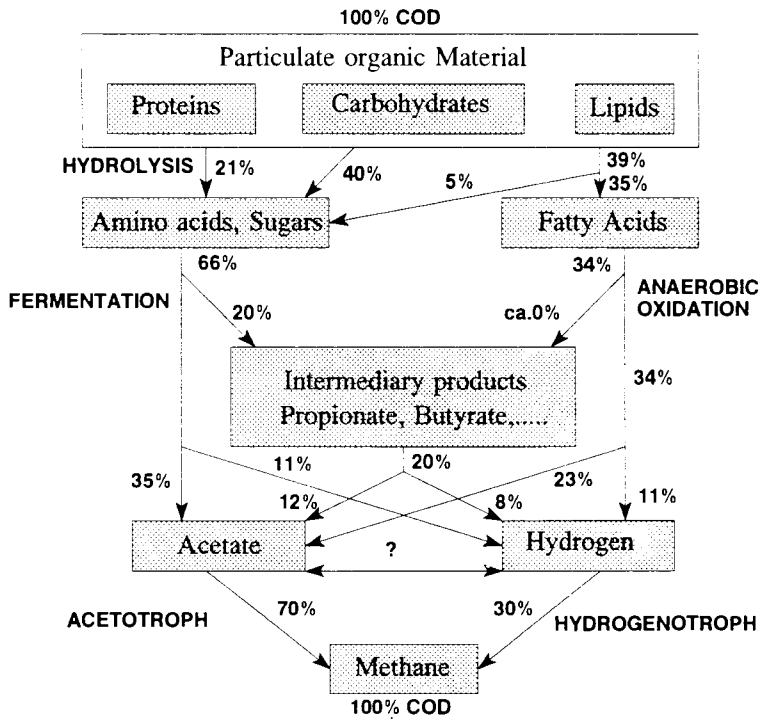

Figure 11a.1 Proposed Reaction Scheme for the Anaerobic Digestion of Domestic Sewage Sludge (Guger and Zehnder, 1983). Percentage Indicates Substrate Flow in the Form of COD Equivalents.

The six distinct conversion processes, considered to be prominent, in the anaerobic digestion process are as follows:

1 The hydrolysis of the biopolymers (proteins, carbohydrates, lipids) into the monomers (amino acids, sugars and long chain fatty acids).

2 The fermentation of amino acids and sugars.

3 The anaerobic oxidation of long chain fatty acids and alchols.

4 The anaerobic oxidation of intermediary products such as volatile fatty acids.

5 The conversion of hydrogen (formate) to methane (hydrogenotrophic methanogenesis).

6 The conversion of actetate to methane (acetotrophic methanogenesis).

11a.3 Setting up the Model

A dynamic model for the anaerobic digestion process, based upon the reaction scheme described by Guger and Zehnder (1983) above, was developed by Siegrist *et al* (1993). The six conversion processes are incorporated into the model with five additional processes included to describe the decay of the five distinct microbial groups. Due to the level of its complexity, the model is not readily manageable and in any case far exceeds the requirements for setting up the current model (to predict COD (VS) destruction and methane production rates). Consequently, in order to provide a simpler model, the rate limiting approach is taken to derive a relatively simple mathematical description of the anaerobic digestion process. The stoichiometry employed by Siegrist *et al* (1993) is however used to describe the fate of the influent COD through the different metabolic pathways (see Table 11a.1 below).

The Rate Limiting Step

For a process composed of a sequence of several reactions, one step is usually much slower than the others and can therefore considered to be the rate limiting step. For the purposes of this particular model, it is accepted that the conversion of acetate to methane (acetotrophic methanogenesis) is the rate limiting step. The methanogenic bacteria which mediate this reaction are notoriously slow growing and the kinetics of their growth dominate the overall rate of reaction.

The rate of hydrolysis of the particulate organics is not considered to effect the overall rate due to the high degree of hydrolysis which takes place in the aerobic reactor; this is one of the major conditioning effects provided by aerobic pre-treatment (see Section 1.3.2) , which reduces the minimum required anaerobic digester retention time.

The Grau Steady State Model

The rate of conversion of acetate to methane is predicted using equation Eq 11a.1 taken from the steady state model of Grau *et al* (1975). For a CSTR (completely stirred tank reactor) at steady state, the effluent substrate concentration S is given by:

$$S = \frac{S_0(1 + b.R_h)}{\hat{\mu}.R_h} \quad \text{...kg(COD)/m}^3 \quad \text{(11a.1)}$$

where:

S = Effluent substrate concentration (kg(COD)/m^3)

S_0 = Influent substrate concentration (kg(COD)/m^3)

$\hat{\mu}$ = Maximum specific growth rate of the organisms (d^{-1})

b = Decay rate of the organisms (d^{-1})

R_h = Solids retention time (d)

The usefulness of the Grau *et al* (1975) equation is that the predicted effluent substrate concentration (S) is a function of the influent substrate concentration (S_0), and therefore is able to take into account the organic loading rate, which has been found to affect digester performance. Pavlostathis and Giraldo-Gomez (1991) consider this an improvement over the Monod equation, where S is independent of S_0.

The Effect of Temperature on the Reaction Rate

The effect of temperature on the reaction rate is predicted by incorporating a temperature variation into the growth and decay rate parameters $\hat{\mu}$ and b. The temperature reltaionships employed are as taken from Buhr and Andrews (1976), who studied both mesophilic and thermophilic anaerobic digestion of sewage sludge:

$$\hat{\mu}_T = 0.324\exp(0.06(T-35°)) \quad .../\text{d} \quad \text{(11a.2)}$$

$$b_T = 0.020\exp(0.14(T-35°)) \quad .../\text{d} \quad \text{(11a.3)}$$

The resulting curve of net growth rate μ_T (= $\hat{\mu}_T$ - b_T) is shown in Figure 11a.2 below, and exhibits a maximum at about 60°C: the optimum operating temperature in the thermophilic region.

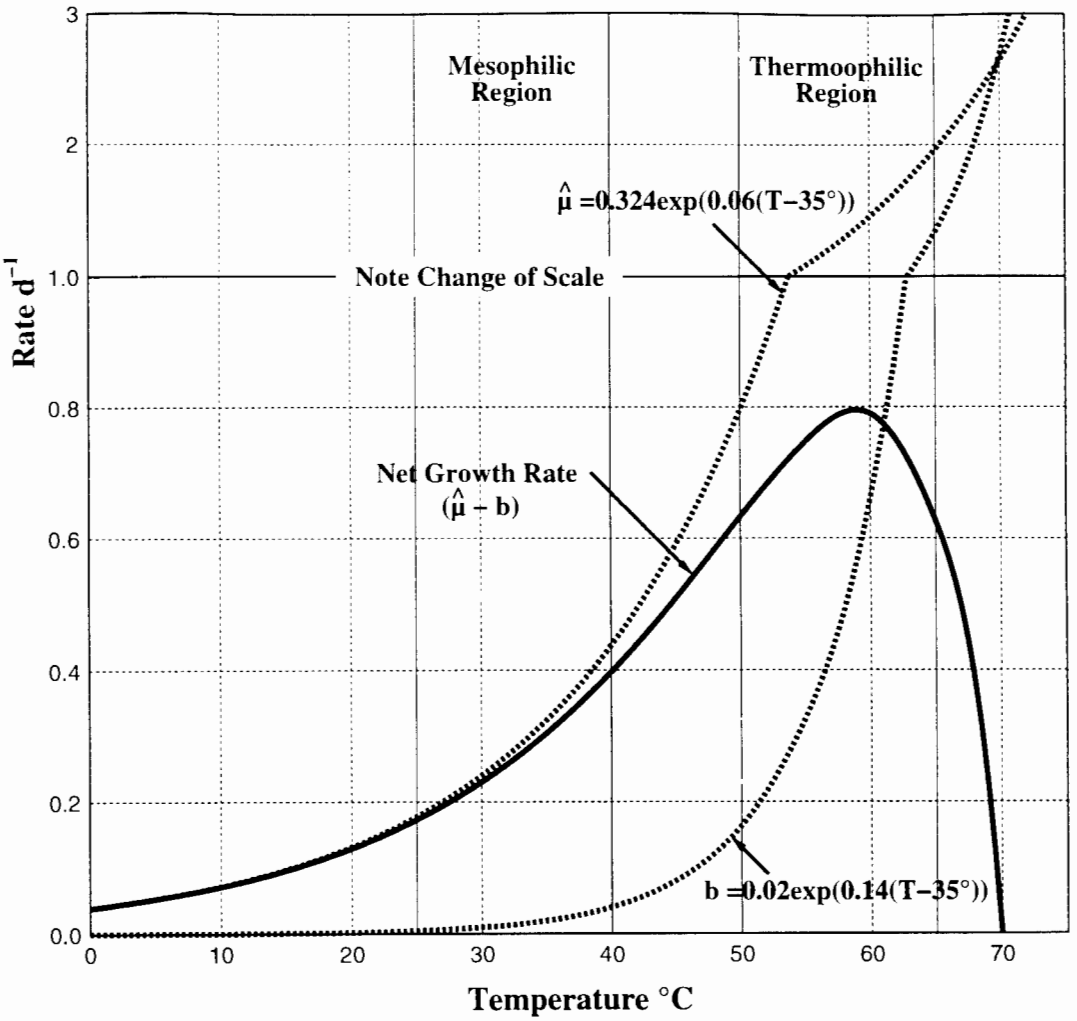

Figure 11a.2 Variation of Growth Rate of Methanogenic Bacteria with Temperature

The Effect of pH on the Reaction Rate

For simplicity, it is presumed that the methanogenic bacteria are not inhibited by the fluctuation or lowering of the pH of the sludge liquor. i.e. the digester has sufficient buffer capacity to sustain the pH within the 7.0 -7.5 pH range. This assumption is considered reasonable for operation at retention times in excess of 6 days.

Process Stoichiometry

The stoichiometry of the digestion process (excluding acetate conversion to methane), obtained from the dynamic model by Siegrist *et al* (1993), is described in Table 11a.1. Starting with 1g(COD) of influent particulate organic matter, the mass of COD of each of the intermediate products in the anaerobic process are defined. The decay of bacterial mass is ignored in calculating the stoichiometry. The rate of death in anaerobic processes is relatively slow in comparison to the rate of growth (~$1/20^{th}$ to $1/5^{th}$ at mesophilic and thermophilic temperatures). The increased complexity which would be brought about by the inclusion of the 'death' processes is not considered warranted.

Table 11a.1 Accepted Process Stoichiometry for the Steady State Model, Derived from the Dynamic Model of Siegrist *et al* (1993)

Process Description (All units reported as mgCOD/gCOD except S_{CO2} reported as mmolCO_2/gCOD)	Particulate Organic Matter X_i							Dissolved Components S_i							Biogas		
		Biomass Generated							Biodegradable Substrate								
	Biodegradable Particulates	Amino acids and sugars →	Fatty acids →	Propionate, Butyrate... →	Hydrogen →	Acetate →	Inert Matter	Amino acids and sugars	Fatty acids	Propionate, Butyrate...	Hydrogen	Acetate	Methane	Carbon Dioxide			
Concentration Symbol →	X_s	X_{as}	X_{fa}	X_{pro}	X_{H2}	X_{ac}	S_i	S_{as}	S_{fa}	S_{pro}	S_{H2}	S_{ac}	S_{CH4}	S_{CO2}			
Input of Biodegradable Particulate COD X_s	**1000**																
Hydrolysis of the Biopolymers X_s	-1000						50	500	450								
Fermention of Amino Acids and Sugars X_{as}		75						-500		106	72	247		6.15			
Anaerobic Oxidation of Fatty Acids X_{fa}			20						-450		137	293		4.07			
Anaerobic Oxidation of Propionate, Butyrate X_{pro}				5						-106	44	57		0.87			
Conversion of Hydrogen X_{H2} to Methane					12						-253		241	-3.3			
Generation of Products Prior to Metanogenesis	**0**	**75**	**20**	**5**	**12**		**50**	**0**	**0**	**0**	**0**	**597**	**241**	**7.79**	Total		**1000**

The generation of Acetate Biomass X_{as} is considered to be the rate limiting step

All the reactions listed in Table 11a.1 are assumed to go to completion because none of them are rate limiting. The objective of Table 11a.1 is to set out the precurser reactions of acetate conversion to methane so that the acetate concentration generated from the 1g(COD) of biodegradable particulates (primary sludge) can be estimated. For this reason only the stoichiometry and not the kinetic relationships of the processes defined in Table 11a.1 are given. Only the rate limiting step is defined kinetically i.e. the acetate to methane conversion rate. The kinetics of this conversion is described below, and for this the Grau Model is accepted.

Application of the Grau Model to Describe Acetogenic Methanogenesis

From the stoichiometry developed from the dynamic model by Siegrist *et al* (1991) and shown in Table 11a.1 above, it can be seen that 1g(COD) of influent biodegradable COD will generate 597mg(COD) of acetate prior to acetogenic mthanogenesis (defined as the acetate yield coefficient Y^*_{ac} = 0.597kg(COD)/kg(COD)).

The mass of acetate remaining (per g influent biodegradable COD) after acetogenic methanogenesis is obtained from the application of the Grau Model (see Eq 11a.1, Grau *et al*, 1975). viz

$$\frac{S_{ac}}{X_s^0} = \frac{Y^*_{ac}(1 + b_T.R_h)}{\hat{\mu}_T.R_h} = \frac{0.597(1 + b_T.R_h)}{\hat{\mu}_T.R_h} \quad \text{...g(COD)/g(COD)} \quad (11a.4)$$

By conducting an acetate mass balance across the system, the mass of acetate destroyed by acetogenic methanogenesis is given by:

$$\frac{Y^*_{ac}.X_s^0 - S_{ac}}{X_b^0} = Y^*_{ac}\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) = 0.597 \times \left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$$

...g(COD)/g(COD) (11a.5)

The amount of 'acetate' biomass generated per g(COD) acetate converted was estimated by Siegrist *et al* (1991) to be Y_{ac}=0.025g(COD)/g(COD). This value is accepted in the model of the acetate to methane rate limiting step. Accordingly, from the conversion of 0.597g(COD) of acetate (which was formed from the input of 1g(COD) of biodegradable substrate), the amount of acetate biomass generated is given by:

$$\frac{X_{ac}}{X_s^0} = Y_{ac}.Y^*_{ac}\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) = 0.0149 \times \left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$$

...g(COD)/g(COD) (11a.6)

The amount of methane generated during the rate limiting acetogenic methanogenesis step, from the conversion of 0.597g(COD) of acetate, is determined from the COD balance for the sub-process:

COD of methane generated = COD of acetate destroyed - COD of biomass generated

Note that in this COD balance, the generation of carbon dioxide (CO_2) and water (H_2O) need not be considered because CO_2 and H_2O are the datum for the COD test (both have zero COD). The rate of carbon dioxide generation during acetogenic methanogenesis will be determined from the carbon balance for the sub-process.

Substitution of Equ's 11a.5 and 11a.6 into the COD balance yields:

$$\frac{S_{CH4}}{X_s^0} = (1 - Y_{ac}).Y_{ac}^{*}\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) = 0.582 \times \left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$$

...g(COD)/g(COD) (11a.7)

The number of moles of methane per g(COD) of methane (equal to the number of moles of carbon in the methane) is given by Siegrist *et al* (1991) as 0.0156 (defined as $i_{C.S_{CH4}}$). Accordingly, the number of moles of methane (equal to moles C in the methane) generated per g(COD) of influent biodegradable COD during acetogenic methanogenesis is given by:

$$\frac{i_{C.X_{CH4}}.S_{CH4}}{X_s^0} = i_{C.X_{CH4}}.(1 - Y_{ac}).Y_{ac}^{*}\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) = 9.079 \times 10^{-3} \times \left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$$

...mol(C)/g(COD) (11a.8)

The amount of carbon dioxide generated during acetogenic methanogenesis is determined by conducting a carbon mass balance across this reaction. i.e.

Carbon content of acetate destroyed	=	Carbon content of biomass generated	+	Carbon content of methane generated	+	Carbon content of CO_2 generated

The stoichiometry of the carbon balance was calculated by Siegrist *et al* (1991), who determined that -0.007 mol(CO_2) were produced per g(COD) biomass generated (defined as $i_{C.X_{as}}$). The fact that the figure is negative shows that carbon dioxide is consumed during acetogenic methanogenesis, with the CO_2 acting as an additional carbon source. Accordingly, the molar mass of carbon dioxide (equal to the molar mass of carbon in the carbon dioxide) consumed per 0.597g(COD) acetate (formed from 1g(COD) influent

biodegradable substrate) can be determined by linking it to Eq 11a.6, which describes the rate of 'acetate' biomass generation. viz:

$$\frac{S_{CO2}}{X_s^0} = i_{CX_{as}}.Y_{ac}.Y_{ac}^{*}\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) = 1.043\times10^{-4}\times\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$$

...mol(C)/g(COD) (11a.9)

All the required stoichiometric coefficients for the non rate limiting reactions leading to the production of acetate, and the kinetic rate equations for the conversion of acetate to products 'acetate' biomass, 'acetate' methane and 'acetate' carbon dioxide are defined. Therefore in the anaerobic process, acetogenic methanogenesis was considered to be the rate limiting step, and the Grau model was utilised to set up the appropriate relationships for the principal components in this reaction. The stoichiometry of the preceeding steps in the anaerobic process were derived from the dynamic model of Siegrist *et al* (1991) and are displayed in Table 11a.1 above. The compilation of the working steady state model is presented below.

11a.4 The Steady State Model

For an anaerobic digester, of process volume V_p (m^3), receiving a daily sludge flow rate of Q_i (m^3/d), the retention time[1] for the process is given by:

$$R_h = \frac{V_p}{Q_i} \quad ...d \quad (11a.10)$$

The Influent Sludge Quality

The organic matter in the influent sludge can be considered to consist of two parts; the biodegradable particulate organic solids X_s^0 ($kg(COD)/m^3$) and the unbiodegradable (inert) organic solids components X_i^0 ($kg(COD)/m^3$). The total influent sludge COD X^0 ($kg(COD)/m^3$) is therefore given by:

$$X^0 = X_s^0 + X_i^0 \quad ...kg(COD)/m^3 \quad (11a.11)$$

At an influent sludge flow rate Q_i (m^3/d), the organic loading rate ***OLR*** to the digester (in terms of COD) is given by:

$$OLR = X^0.\frac{Q_i}{V_p} = \frac{X^0}{R_h} = \frac{X_s^0 + X_i^0}{R_h^0} \quad ...kg(COD)/m^3.d \quad (11a.12)$$

---

[1] With no supernatant withdrawal, the hydraulic and solids retention time are equivalent.

The Anaerobic Digestion Process

It is accepted that all the influent biodegradable particulate organic matter X_s^0 undergoes enzyme hydrolysis. The resulting soluble substrate passes through a number of conversion processes, producing various types of biomass, intermediate and end products (see Figure 11a.1 above).

Accepting that the influent sludge contains no anaerobic biomass or acetate, then the following mass balance applies to each of the components (*j*) generated in the process:

Rate of Generation of Component j = Rate of Output of Component j

The generation rate (GR_j) of component j is determined from the stoichiometric coefficients or formulae (abbreviated as a_j) developed in Section 11a.3 above:

$$GR_j = a_j.LR_{X_s^0} = a_j.\frac{X_s^0}{R_h} \quad \text{...kg(COD)/m}^3\text{.d} \quad \text{(11a.13)}$$

For the generation rates of the biogas components methane and carbon dioxide, the stoichiometric formulae a_j are as mol(C)/g(COD). The concentration of methane in the biogas $\%CH4$ is obtained by dividing the rate of methane production GR_{CH4} with the total rate of biogas production $GR_{CH4} + GR_{CO2}$ viz:

$$\%CH4 = 100.\frac{GR_{CH4}}{GR_{CH4}+GR_{CO2}} = 100.\frac{a_{CH4}}{a_{CH4}+a_{CO2}} \quad \text{...\%} \quad \text{(11a.14)}$$

where a_{CH4} and a_{CO2} are the formulae for the production of methane and carbon dixide respectively.

The Effluent Sludge Quality

The output rate OR_j of component j, which has been generated in the anaerobic process, is determined from the product of the flow rate through the process Q_i per unit process volume V_p and the concentration C_j of the component in the effluent sludge: viz

$$OR_j = C_j.\frac{Q_i}{V_p} = \frac{C_j}{R_h} \quad \text{...kg(COD)/m}^3\text{.d} \quad \text{(11a.15)}$$

Accordingly, the concentration (C_j) of component j in the effluent sludge is given by:

$$C_j = a_j.X_s^0 \quad \text{...kg(COD)/m}^3 \quad \text{(11a.16)}$$

The unbiodegradable particulate organic matter X_i^0 which enters the digester with the influent sludge, is assumed to pass through the process unchanged i.e.

$$X_i = X_i^0 \quad \text{...kg(COD)/m}^3 \quad \text{(11a.17)}$$

The total effluent sludge COD X is therefore given by:

$$X = X_i^0 + \sum a_j.X_s^0 \quad \text{...kg(COD)/m}^3 \quad \text{(11a.18)}$$

The total output rate of COD from the digester OR is given by

$$OR = \frac{X_i^0}{R_h} + \frac{\sum a_j.X_s^0}{R_h} \quad \text{...kg(COD)/m}^3\text{.h} \quad \text{(11a.19)}$$

The Rate of COD Destruction in the digestion process COD_{dest} is obtained from the difference between the organic loading rate OLR (Eq11a.12) and the output rate OR (Eq11a.19) viz:

$$COD_{dest} = \frac{X_s^0 + X_i^0}{R_h} - \frac{X_i^0 + \sum a_j.X_s^0}{R_h} = \frac{X_s^0 - \sum a_j.X_s^0}{R_h} \quad \text{...kg(COD)/m}^3\text{.h} \quad \text{(11a.20)}$$

The COD removal efficiency (or %COD removal) RE_{COD} is defined as the ratio of the COD destruction rate COD_{dest} (11a.20) and the organic loading rate OLR (Eq 11a.12) expressed as a percentage viz:

$$RE_{COD} = 100.\frac{(COD_{dest})}{(OLR)} = 100.\frac{\left(X_s^0 - \sum a_j.X_s^0\right)}{\left(X_s^0 + X_i^0\right)} \quad \text{...kg(COD)/m}^3\text{.h} \quad \text{(11a.21)}$$

Formulae for the Steady State Model

Formulae for the fundamental parameters of the steady state model are listed in tabular form below (Tables 11a.2 to 11a.4). The principal equations generated in the Tables are listed at the end of the text.

Table 11a.2 Influent Digester Sludge Quality for the Steady State Model

Component	Concentration (kg(COD)/m^3)	Reference
Biodegradable Particulates	X_s^0	-
Non-biodegradable Particulates	X_i^0	-
Total Influent COD	$X_s^0 + X_i^0$	Eq (11a.11)

Table 11a.3 Effluent Digester Sludge Quality for the Steady State Model

Component	Concentration (kg(COD)/m^3)	Reference
'Amino acid' Biomass (X_{as})	$0.075 \times X_s^0$	Table 11a.1
'Fatty Acid' Biomass (X_{fa})	$0.020 \times X_s^0$	" "
'Propionate' Biomass (X_{pro})	$0.005 \times X_s^0$	" "
'Hydrogen' Biomass (X_{H2})	$0.012 \times X_s^0$	" "
'Acetate' Biomass (X_{ac})	$0.0149 \times \left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) . X_s^0$	Eq (11a.6)
Inert Particulate Matter (X_i)	X_i^0	-
Inert Soluble Matter (S_i)	$0.050 \times X_s^0$	Table 11a.1
Biodegradable Soluble Matter (S_{as})	$0.597 \times \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h} . X_s^0$	Eq (11a.4)
Total Effluent COD	$X_i^0 + 0.177X_s^0 + 0.582\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h} . X_s^0$	Eq (11a.18)

Table 11a.4 Volatile Solids Loading Rate and Methane Generation Rate

Component	Rate (kg(COD)/m^3.d)	Reference
Volatile Solids (COD) Loading Rate to the Digester	$\frac{X_s^0 + X_i^0}{R_h}$	Eq(11a.12)
Volatile Solids (COD) Destruction Rate inside the Digester	$\frac{X_s^0}{R_h}\left(0.823 - 0.582\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$	Eq(11a.20)
Methane Generation Rate: Prior to Acetogenic Methanogenesis	$0.241 \times X_s^0$	Table 11a.1
Methane Generation Rate During Acetogenic Methanogenesis	$0.582\left(1 - \frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right) . X_s^0$	Eq (11a.7)
Total Methane Generation Rate	$\frac{X_s^0}{R_h}\left(0.823 - 0.582\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)$	-
%COD Destruction inside the Digester	$\frac{X_s^0.\left(82.3 - 58.2\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}\right)}{\left(X_s^0 + X_i^0\right)}$	Eq (11a.21)

Note that the rate of COD destruction is identical to the rate of methane generation (measured in terms of COD). This is to ensure COD continuity across the process. The rate of methane production in molar mass units is presented in Table 11a.5 below.

Table 11a.5 Biogas Production Rates and Quality in Terms of Methane Concentration

Component	Rate (mol(C)/m^3.d)	Reference
Methane Generation Rate	$12.83\frac{X_s^0}{R_h} - 9.08\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h} \cdot \frac{X_s^0}{R_h}$	-
CO_2 Generation Rate	$7.69\frac{X_s^0}{R_h} + 0.104\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h} \cdot \frac{X_s^0}{R_h}$	-
Total Biogas Generation Rate	$20.52\frac{X_s^0}{R_h} - 8.98\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h} \cdot \frac{X_s^0}{R_h}$	-
Biogas Methane Concentration %	$100\% \times \left(\frac{12.83 - 9.08\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}}{20.52 - 8.98\frac{(1 + b_T.R_h)}{\hat{\mu}_T.R_h}} \right)$	Eq (11a.14)

Graphical Demonstration of the Model

A graphical demonstration of the steady state model is presented in Figures 11a.3 to 11a.5 below. The Figures show the change in:

- Percent volatile solids destroyed (%COD/COD) Figure 11a.3
- Biogas Production Rate (mol/m^3.d) Figure 11a.4
- Methane concentration in the biogas (%) Figure 11a.5

with the **Solids Retention Time** (d) at different **Operating Temperatures** (°C). The influent COD is taken to be 45kg(COD)/m^3 with a biodegradable fraction of 60%.

Note that the model is not valid at retention times less than 5 days. It is considered that at short retention times, the anaerobic processes prior to acetotrophic methanogenesis will influence the overall reaction rate, and that the pH cannot be assumed to be in the optimum range for methanogenesis; The simplifications made in the model (see Section 11a.3) therefore do not apply at retention times less than 5 days.

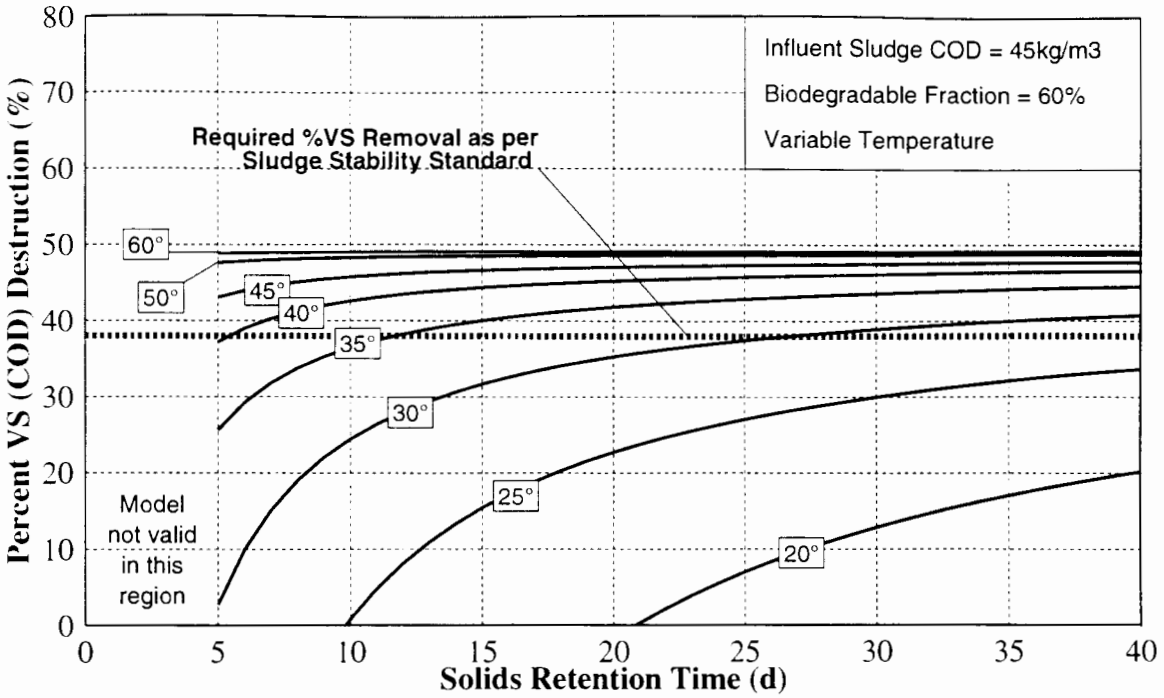

Figure 11a.3 The Change in Percent VS (COD) Destruction with Solids Retention Time at Different Operating Temperatures: The Influent COD equals 45kg(COD)/m^3 with the Biodegradable Fraction at 60%.

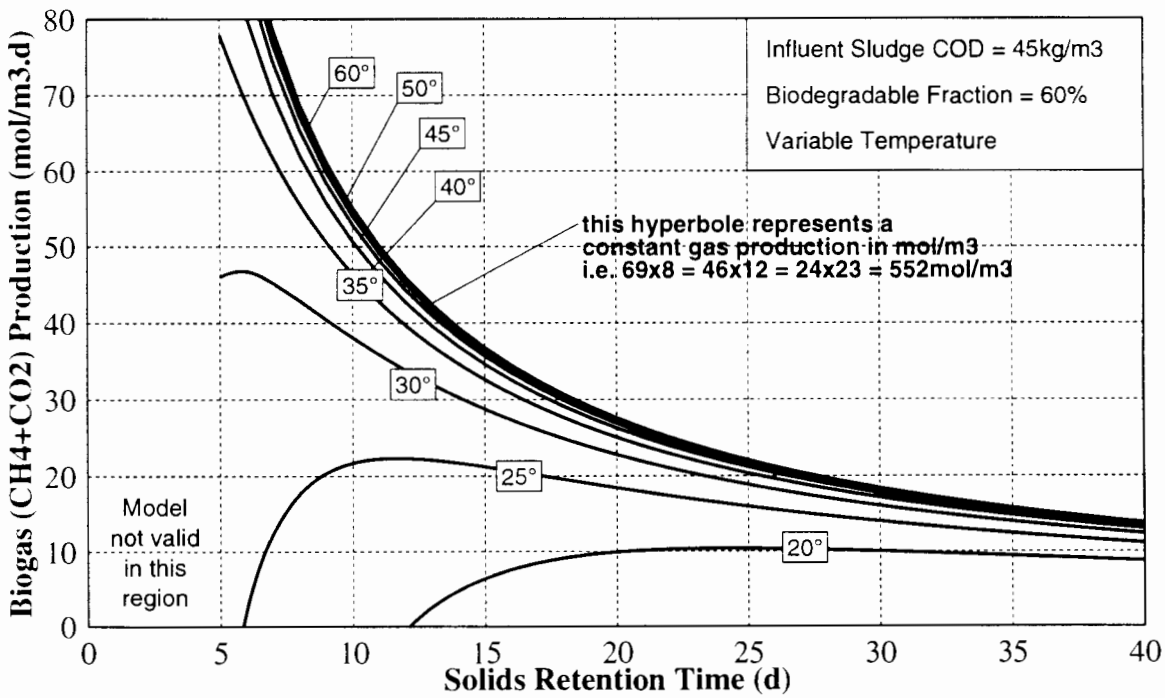

Figure 11a.4 The Change in Biogas Production (molC/m^3.d) with Solids Retention Time at Different Operating Temperatures: The Influent COD equals 45kg(COD)/m^3 with the Biodegradable Fraction at 60%.

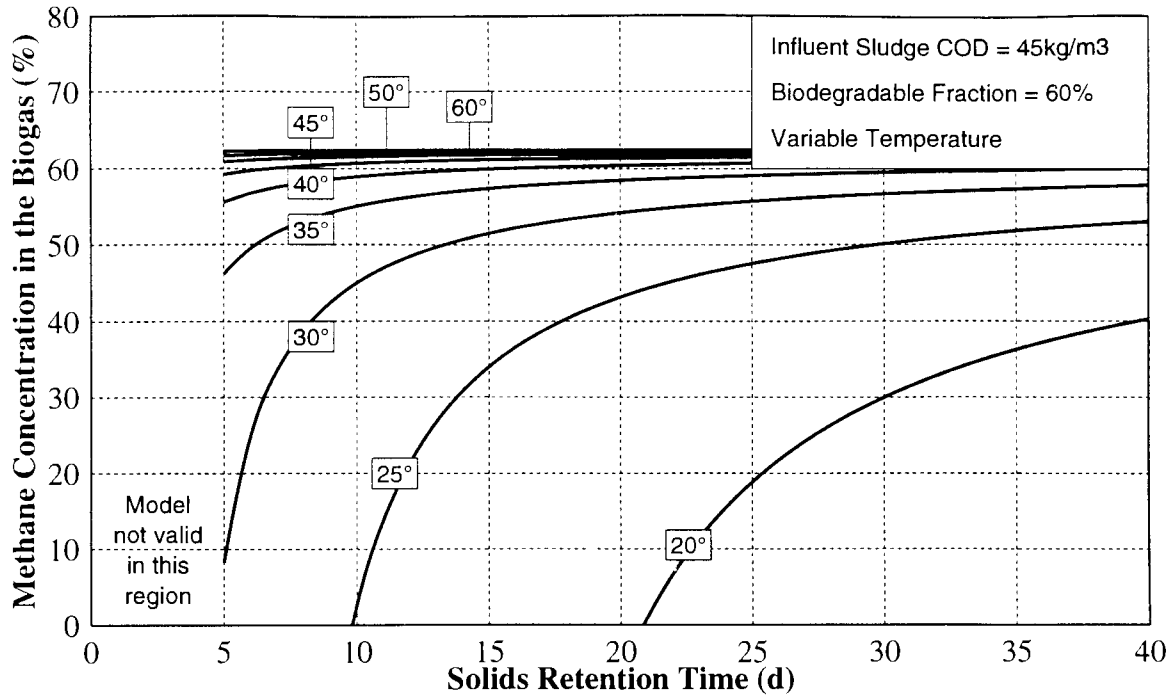

Figure 11a.5 The Change in Biogas Methane Concentration (%) with Solids Retention Time at Different Operating Temperatures: The Influent COD equals 45kg(COD)/m^3 with the Biodegradable Fraction at 60%.

At a fixed **retention time of 10 days**, Figure 11a.6 below shows the variation in COD destruction, biogas production, and methane concentration with digester **temperature**. This graphically illustrates the difficulty in achieving satisfactory digestion at mesophilic temperatures at retention times around 10 days. In contrast, at the same retention time, at thermophilic temperatures the process is more stable and capable of satisfactory performance.

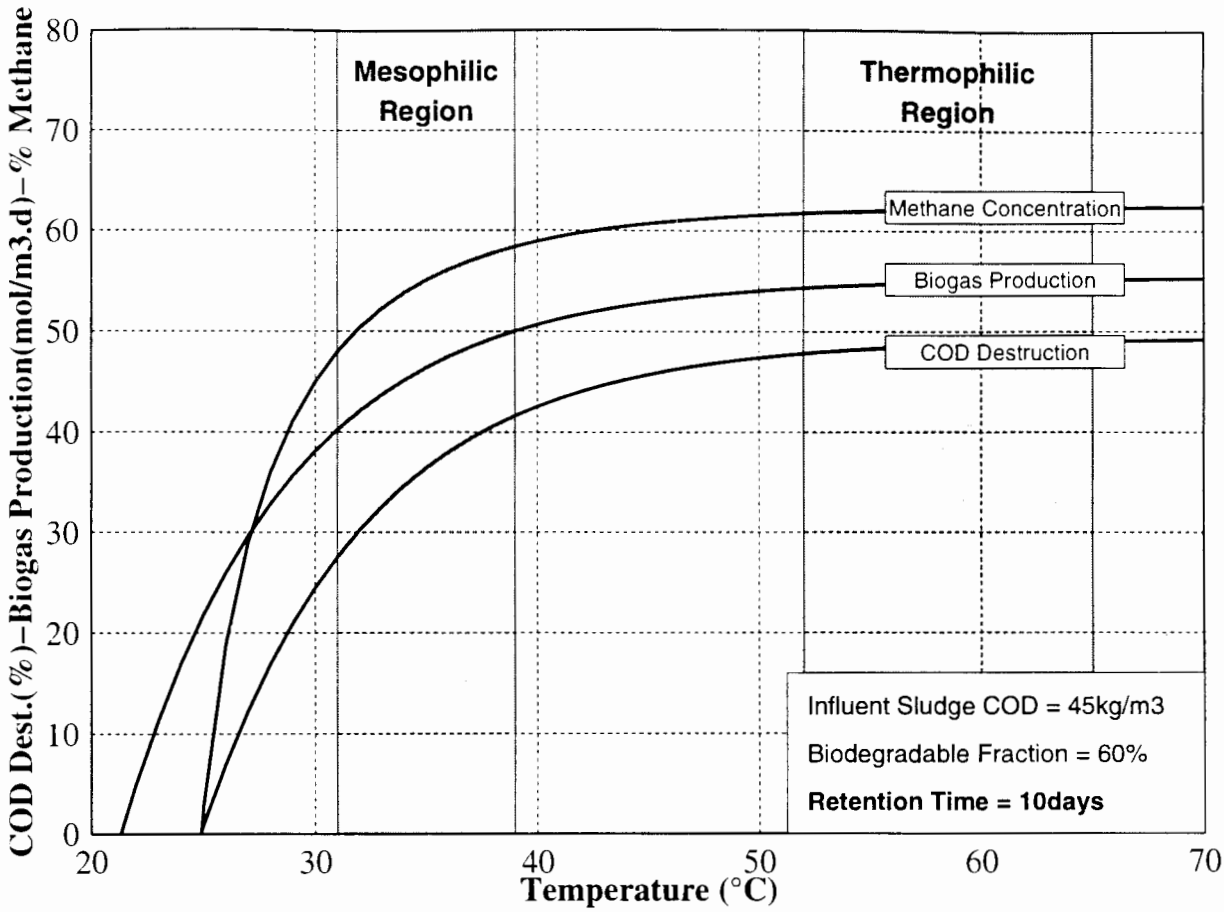

Figure 11a.6 The Change in COD Destruction (%), Biogas Production (molC/m^3.d), and Biogas Methane Concentration (%) with Digester Temperature at a Fixed Solids Retention Time of 10 days: The Influent COD equals 45kg(COD)/m^3 with the Biodegradable Fraction at 60%.

11a.5 Modelling The *SOUR* Final Sludge Stability Test

In the past, emphasis has all too often be placed on operating the digestion process at the shortest possible retention time (in order to increase treatment capacity), without taking cognisance of the stability of the final sludge product. It is a fact that the anaerobic process itself is stable at lower retention times than that required to stabilise the sludge. It is therefore considered more appropriate to determine the minimum sludge retention time in terms of the quality of the product than the stability of the anaerobic process.

The two standard measurements of final sludge stability are (1) a %VS removal of >38% and (2) a specific oxygen utilisation rate ***SOUR*** < 1.0 g(O_2)/kg(TSS).h (Heidman, 1989). The difficulty of applying these two standards to the final sludge quality predicted by the anaerobic process model described above, is that whilst it is relatively easy for the process to achieve a >38% reduction when treating primary sludge (%TS>4%) the

resulting ***SOUR*** will not necessarily meet with the <1.0 $g(O_2)/kg(TSS).h$ standard[2]. Consequently the following criterion is used in conjunction with the model:

To be classified as stable, the quality of the final sludge must be equal to or better than that of primary sludge (4%TS) after 20 days of conventional mesophilic digestion at 35°C. The final sludge quality, as predicted by the process model is given in Table 11a.6 below.

Table 11a.6 The Anaerobic Digestion of Primary Sludge (R_h=20 days): Predicted Influent and Effluent Sludge Quality.

Parameter		Influent	Effluent	% Removal
Total Solids	kg/m^3	40.0	26.4	34.0
Volatile Solids	kg/m^3	32.4	18.8	42.0
Biodegradable Volatile Solids	kg/m^3	19.4	2.5	87.1

The criterion then simplifies to "the final sludge must have a biodegradable volatile solids concentration < 2.5 $kg(BVS)/m^3$ " to qualify as a stable sludge. It is this condition which is placed in the general dual digestion simulation model described in Chapter 8, in the determination of the minimum retention time for satisfactory digestion (refer Section 8.3.6).

In an attempt to simulate the ***SOUR*** test described in Appendix 7, an aerobic kinetic model was developed (based on the kinetics and stoichiometry employed by Dold *et al*, 1991). The acetate concentration (RBVS) remaining after anaerobic digestion is accepted to be the available source of substrate for the growth of aerobic bacteria in a non hydrolysed form i.e. as slowly biodegradable particulated COD (SBCOD). Values for the initial aerobic biomass concentration and maximum specific growth rate on SBCOD were adjusted until the simulated ***SOUR*** values matched the actual test data (see Figure 11a.7 below). The principle functions in the model are:

---

[2] At the end of phase II, a sample of final (effluent) sludge from one of the conventional anaerobic digesters at Athlone (20 day retention time) was tested for ***SOUR*** (see Appendix 7). Whilst the %VS removal achieved in the process was in excess of 50%, in the ***SOUR*** test a peak of 13 $g(O_2)/kg(TSS).h$ was produced during the first 24 hours. At no point in the ***SOUR*** test which lasted 96 hours, after the growth of an adequate population of aerobic bacteria, did the ***SOUR*** reach the target value of <1 $g(O_2)/kg(TSS).h$ (see Figure 11a.7).

1: **The Utilisation of Biodegradable Substrate (COD)**

$$\frac{dS_{bs}}{dt} = \frac{-1}{Y_{ZH}}.K_{MP}.\frac{S_{bs}/Z_{BH}}{K_{SP} + S_{bs}/Z_{BH}}.Z_{BH} + (1 - f_e).b_h.Z_{BH} \quad \text{...kg(COD)/m}^3\text{.h} \quad (11a.22)$$

2: **The Growth (and Decay) of Aerobic Bacteria**

$$\frac{dZ_{BH}}{dt} = K_{MP}.\frac{S_{bs}/Z_{BH}}{K_{SP} + S_{bs}/Z_{BH}}.Z_{BH} - b_h.Z_{BH} \quad \text{...kg(COD)/m}^3\text{.h} \quad (11a.23)$$

2: **The Oxygen Utilisation Rate**

$$\frac{dO}{dt} = 1000.\frac{1 - Y_{ZH}}{Y_{ZH}}.K_{MP}.\frac{S_{bs}/Z_{BH}}{K_{SP} + S_{bs}/Z_{BH}}.Z_{BH} \quad \text{...g(O}_2\text{)/m}^3\text{.h} \quad (11a.24)$$

where:

S_{bs} = biodegradable volatile solids concentration assumed SBCOD (kg(COD)/m^3)

Z_{BH} = heterotroph active biomass concentration (kg(COD)/m^3)

dO/dt = oxygen utilisation rate (g(O$_2$)/m^3.h)

Y_{ZH} = heterotrophic biomass yield (kg(COD)/kg(COD)

K_{MP} = heterotroph maximum specific growth rate on SBCOD (=0.25/h)

K_{SP} = half saturation constant for SBCOD (= 0.027 kg(COD)/kg(COD))

f_e = endogenous residue fraction (= 0.20 kg(COD)/kg(COD))

b_h = endogenous respiration rate of the heterotrophic biomass (=0.010 /h)

The above equations are solved by integrating stepwise forward in time starting at time t=0. The initial heterotrophic active biomass concentration was taken as 0.08 kgCOD/m^3 and the maximum specific growth rate as K_{MP} = 0.25/h. Both of the values were adjusted to match the estimated peak with the actual peak. The computer programme for determining the ***SOUR*** values for the primary sludge (4%TS) after 20 days digestion (described in Table 11a.6) is listed at the end of this Appendix. A similar programme has been incorporated into the general dual digestion simulation model described in Chapter 8. In this general model, the predicted peak value for the ***SOUR*** (g(O$_2$)/kg(TSS).h) for the final (digester effluent) sludge product is given together with the time at which it occurs. For the primary sludge (4%TS) after 20d digestion the peak ***SOUR*** value is calculated at 10.7 g(O$_2$)/kg(TSS).h and occurs after 14.5 hrs (Figure 11a.7).

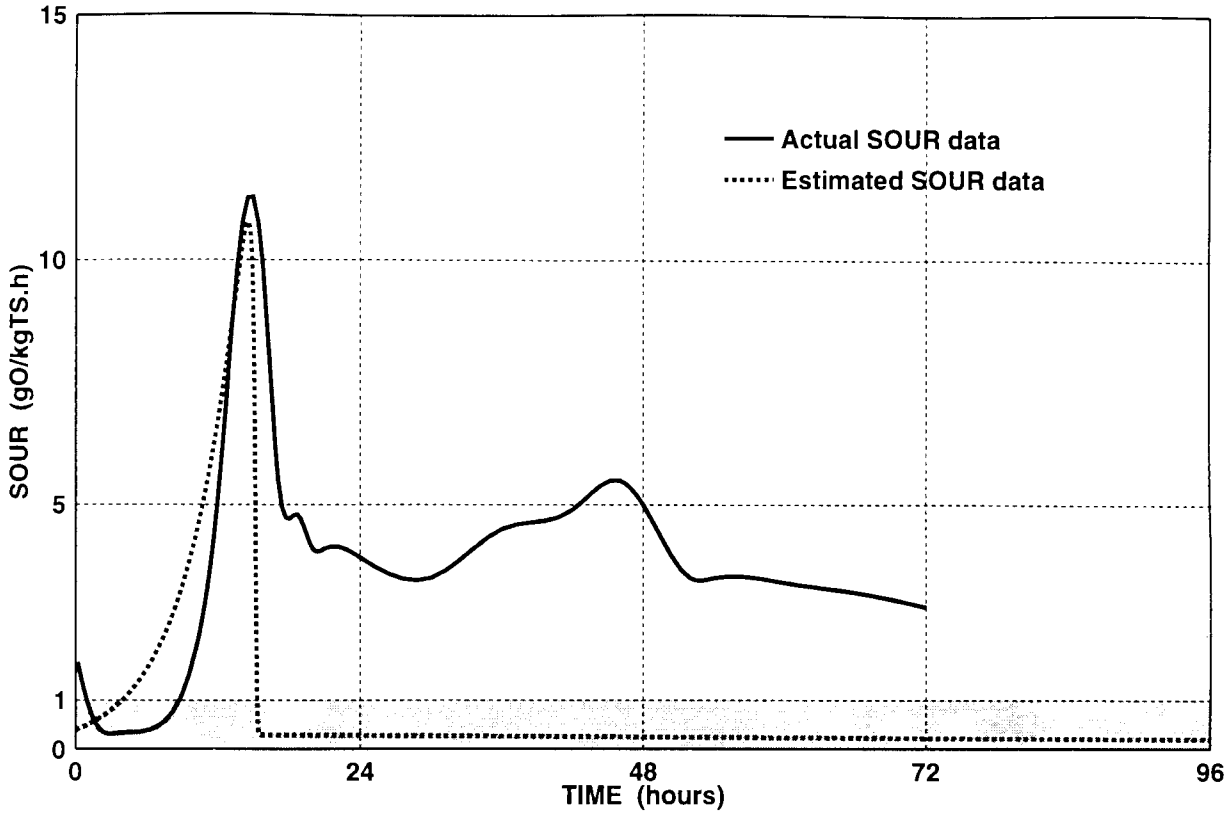

Figure 11a.7 Plot of Predicted and Actual *SOUR* Data for Anaerobically Digested Sludge (20 day Retention Time at 35°C).

Listing of Computer Programme

```
#include <stdio.h>
double sqrt();
double pow();
main()
{
sourprog();
}
/***SIMULATION OF THE SOUR TEST***/
/***/sourprog()
{
double Sa,Sa1,Sai,Zbh,Zbh1,Zbhi,Ksp,Kmp,Yzh,time,timemax,deltat;
double dSadt,dOdt,dZbhdt,bh,fe;
double fovs,CTSd,CNBVSi,CVSd,SOUR,SOURmax;
fovs   = 1.700;
CTSd   = 26.400;
CNBVSi = 16.300;
CVSd   = 18.800;
Sai    = fovs*(CVSd-CNBVSi);
Kmp = 0.25;          /***kgCOD/kgCOD-cells.h***/
Yzh = 0.660;         /***kgCOD/kgCOD-cells***/
```

```
Ksp = 0.027;          /***kgCOD/kgCOD***/
Zbhi = 0.08;          /***kgCOD/m3****/
deltat = 0.01;        /****h****/
fe  = 0.200;          /***kgCOD/kgCOD***/
bh  = 0.01;           /*** /h***/
Zbh = Zbhi;
Sa  = Sai;
time = 0.000;
do
{
  dSadt  = (-1.000/Yzh)*Kmp*((Sa/Zbh)/(Ksp+(Sa/Zbh)))*Zbh + (1.000-fe)*bh*Zbh;
  dZbhdt  = (Yzh*(1.000/Yzh)*Kmp*((Sa/Zbh)/(Ksp+(Sa/Zbh)))*Zbh) - (bh*Zbh);
  Sa1   = Sa + (dSadt*deltat);
  Sa    = Sa1;
  Zbh1   = Zbh + (dZbhdt*deltat);
  Zbh    = Zbh1;
  dOdt   = (1000.000*(1.000-Yzh)/Yzh)*Kmp*((Sa/Zbh)/(Ksp+(Sa/Zbh)))*Zbh;
  if ((dOdt/CTSd) > SOUR)
    {
    SOURmax = SOUR;
    timemax = time;
    }
  SOUR   = dOdt/CTSd;
/* printf("%6.3lf  %6.3lf\n",time,SOUR);*/
  time   = time + deltat;
}
while (time <= 96.00);
printf("The Maximum Substrate Oxygen Utilisation Rate     = %6.3lf gO/kgTSS.h\n",SOURmax);
printf("(this Occurred after %6.1lf hrs)\n",timemax);
}
```

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APPENDIX 12

THE ANAEROBIC DIGESTER STEADY STATE HEAT BALANCE

12a.1 The Need to Control Digester Temperature

Presented in this Appendix is a simple model, based on the anaerobic digester steady state heat balance, employed to predict the anaerobic digester operating temperature in the dual digestion process. In dual digestion plants, where no heat exchange facility[1] exists between aerobic reactor and anaerobic digester, the resultant digester operating temperature is dependent upon the operating conditions set up for the aerobic reactor. In South Africa, with the relatively high ambient temperatures encountered during the summer months (~25°C), the digester when operated at retention times <15 days has the potential to reach temperatures in excess of the mesophilic temperature range (>38°C). This potential was realised by Messenger *et al* (1992) on the dual digestion plant at Milnerton. Temperatures in excess of 40°C were encountered in the digester during the summer months when the retention time was reduced to 12 days. At Athlone, during phase II the digester temperature reached thermophilic temperatures (>50°C) during the summer months at a retention time of 10 days (see Section 7.2.4).

It is essential therefore, that if no heat exchange facility is provided between reactor and digester, cognisance must be taken of the resulting anaerobic digester temperature when selecting operating conditions for the aerobic reactor. It is important to ensure that the digester temperature does not fall between the mesophilic (28-38°C) and thermophilic (50-70°C) temperature ranges as this will seriously inhibit the performance of the anaerobic digestion process. If this aspect is considered essential for satisfactory dual digester performance, and considering the requirements for pasteurisation, then in the selection of operating conditions for the aerobic reactor the following two criterion should be applied:

---

[1] Dual digestion plants are frequently fitted with a heat exchanger to transfer sensible heat from the hot sludge discharged from the aerobic reactor to the cold reactor influent feed sludge. The quantity of heat transferred can be controlled by adjusting the retention time in the exchanger. In this manner a certain degree of control can be exercised over the anaerobic digester temperature, ensuring that it operates at optimum mesophilic temperatures.

- The aerobic reactor **must** operate at a temperature in excess of 60°C to ensure satisfactory pasteurisation.

- The flow rate of sludge Q leaving the reactor (discharged to the digester) and the temperature of the reactor sludge T_r must be such that the sensible heat input to the digester H_s ($=C_p.Q.(T_r-T_d)$) is appropriate to maintain the digester temperature T_d in either the mesophilic or thermophilic range.

Indeed, in modelling the dual digestion process (Chapter 8) it is recommended that these two constraints be placed on the process in order to achieve pasteurisation and efficient stabilisation. In order to be able to predict the anaerobic digester temperature, or conversely if the digester temperature is fixed to predict the required aerobic reactor temperature, at different sludge flow rates and ambient temperatures it is necessary to perform a steady state heat balance around the digester.

12a.2 The Anaerobic Digester Steady State Heat Balance

A simplified model for the steady state heat balance around the digester is presented. It is presumed that the sensible heat given up by the sludge through the digestion process is passed by conduction/convection through the walls of the digester to the surroundings. No consideration is given to heat input due to mechanical mixing or vapour heat losses as a result of gas mixing[2]. The anaerobic digester steady state heat balance is therefore as follows:

$$C_p.\frac{Q(SL)_{in}}{24}.(T(SL)_r - T(SL)_d) = U_0^{dig}.A_{dig}.(T(SL)_d - T_{amb}) \quad \text{...MJ/h} \quad (12a.1)$$

where:

C_p = The specific heat capacity of the sludge (≈4.0 MJ/m^3.°C refer Section 3.7.2)
$Q(SL)_{in}$ = The sludge flow rate through the digester (m^3/d)
$T(SL)_r$ = Temperature of the sludge leaving aerobic reactor/entering digester (°C)
$T(SL)_d$ = Temperature of the sludge leaving the anaerobic digester (°C)
U_0^{dig} = Overall heat transfer coefficient (MJ/m^2.h.°C)
A_{dig} = Wall area covered by the hot sludge in the digester (m^2)
T_{amb} = Ambient Temperature (°C)

Equation 12a.1 expressed in terms of retention time takes the form:

---

[2] No metering facilities were available to measure the biogas flow rate into and out of the digester during the Investigation at Athlone (both phase I and II).

$$C_p \cdot \frac{V_{dig}}{24 \cdot R_h} \cdot (T(SL)_r - T(SL)_d) = U_0^{dig} \cdot A_{dig} \cdot (T(SL)_d - T_{amb}) \quad \text{...MJ/h} \quad (12a.2)$$

where:

V_{dig} = The process volume of the digester (m^3)

R_h = The hydraulic retention time in the digester (d)

In order to find a satisfactory solution to Eq 12a.2, an accurate estimate of the overall heat transfer coefficient U_0^{dig} needs to be made.

12a.3 Estimation of the Overall Heat Transfer Coefficient for the Digester

An estimate of the overall heat transfer coefficient U_0^{dig} is made by fitting actual operating data (Table 12a.2) obtained from full scale dual digestion plants (Table 12a.1) to Eq 12a.2 above, rearranged in terms of U_0^{dig} viz:

$$U_0^{dig} = \frac{C_p \cdot V_{dig} \cdot (T(SL)_r - T(SL)_d)}{24 \cdot R_h \cdot A_{dig} \cdot (T(SL)_d - T_{amb})} \quad \text{...MJ/m}^2\text{.h.°C} \quad (12a.3)$$

The dual digestion plants for which data was available are listed in Table 12a.1 below. The periods of operation for which the data applies (Table 12a.2) is listed along with the appropriate references. Digester operating data from three prolonged periods of stable operation at the Athlone Plant are also supplied.

Table 12a.1 Full Scale Dual Digestion Plants for which Data is Obtained to Estimate A Value for the Overall Heat Transfer Coefficient for the Digester. Included are the Periods of Operation and References.

Plant	Period of Operation	Reference
Milnerton, RSA	2 May 89 to 2 August 89	Messenger *et al* (1992)
Hagarstown, USA	23 June 80 to 2 Nov 80	Appleton and Venosa (1986)
Lackawanna, USA	3 July 84 to 10 Sept 84	Appleton and Venosa (1986)
Athlone[1], RSA	5 Oct 89 to 12 Aug 90	Section 4.2.1; Table 4.1
Athlone[2], RSA	21 July 94 to 24 Aug 94	Section 7.2.1; Figure 7.1
Athlone[3], RSA	22 Nov 94 to 14 Dec 94	Section 7.2.1; Figure 7.1

The appropriate operating data obtained from the four plants listed in Table 12a.1 above is presented in Table 12a.2 below:

Table 12a.2 Full Scale Dual Digestion Plant Data for the Estimation of a Value for the Overall Heat Transfer Coefficient for the Digester.

Plant	Process Volume V_{dig} (m^3)	Surface Area A_{dig} (m^2)	Sludge Flow Rate $Q(SL)_r$ (m^3/d)	Retention Time R_h (d)	Sludge Temp In $T(SL)_r$ (°C)	Sludge Temp Out $T(SL)_{dig}$ (°C)	Ambient Temp T_{amb} (°C)	Calculated O.H.T.C. U_0^{dig} ($MJ/m^2.h.°C$)
Milnerton	536	536	29	18.5	58	37.4	16	0.0086
Hagarstown	670	532	50	13.4	50.5	41.7	21.5	0.0069
Lackawanna	440	322	39.4	11.2	52.5	43.6	18.4	0.0072
Athlone[1]	1800	1243	41	42	49	31	20	0.0089
Athlone[2]	1800	1243	96	19	50	37	17	0.0084
Athlone[3]	1800	1243	192	9.4	62	53	23	0.0077

From the calculated values for the overall heat transfer coefficient U_0^{dig} determined from the application of actual full scale data (Table 12a.2) using Eq 12a.3, an average value for U_0^{dig} = 0.008 $MJ/m^2.h.°C$ is accepted for use in Eq 12a.2(1) above. This value for U_0^{dig} is equivalent to 2.2 $W/m^2.°C$ which is in good agreement with the value of 2 $W/m^2.°C$ quoted by Bruce and Oliver (1987) for modern package digesters with full insulation. The steady state heat balance for the digester (Eq 12a.2) can now be expressed as:

Sensible Heat Given Up by the Sludge = **Heat Loss From the Digester**

$$4.0 \times \frac{V_{dig}}{24.R_h}.(T(SL)_r - T(SL)_d) = 0.008 \times A_{dig}.(T(SL)_d - T_{amb}) \quad \text{...MJ/h} \quad (12a.4)$$

The steady state heat balance is depicted graphically in Figure 12a.1 below. Using the full scale data presented in Table 12a.2 above, the sensible heat given up by the sludge is plotted against the heat loss from the digester.

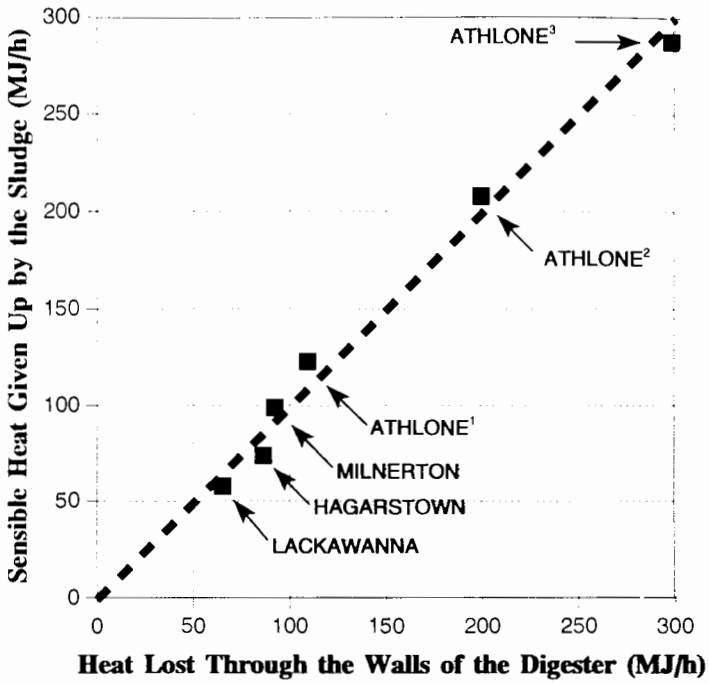


Figure 12a.1 Sensible Heat Given up by the Sludge and the Heat Lost Through the Walls of Digesters (MJ/h) from Selected Full Scale Dual Digestion Plants (as listed in Table 12a.1).

12a.4 Application of the Anaerobic Digester Steady State Heat Balance

With a function now defined (Eq 12a.4) to describe the steady state heat balance for the anaerobic digester in dual digestion systems, three applications are now described which demonstrate its value in providing the link between aerobic reactor and anaerobic digester performance.. Note that Eq 12a.4 plays an integral role in the general model described in Chapter 8 to simulate the dual digestion process.

Application 1: Fixed Aerobic Reactor Temperature (60°C)

In the first application, the aerobic reactor temperature is fixed at 60°C (the temperature required for pasteurisation). Rearranging Eq 12.4 in terms of digester temperature, and substituting for $T(SL)_r$=60°C, yields:

$$T(SL)_d = \frac{(4.0 \times 60 \times V_{dig}) + (0.008 \times 24 \times A_{dig}.T_{amb}.R_h)}{(24 \times 0.008 \times A_{dig}.R_h) + (4.0 \times V_{reac})} \quad ...°C \quad (12a.5)$$

For the Athlone digester V_{dig}= 1800m^3 and A_{dig}= 1243m^2. Substitution into Eq 12a.5 yields:

$$T(SL)_d = \frac{(4.0 \times 60 \times 1800) + (0.008 \times 24 \times 1243 \times T_{amb}.R_h)}{(24 \times 0.008 \times 1243 \times R_h) + (4.0 \times 1800)} \quad ...°C \quad (12a.6)$$

which upon simplification yields:

$$T(SL)_d = \frac{1810 + T_{amb}.R_h}{R_h + 30.2} \quad \ldots°C \quad (12a.7)$$

Eq 12a.7 is depicted graphically in Figure 12a.2 below, for ambient temperatures T_{amb} ranging from 0° to 25°C.

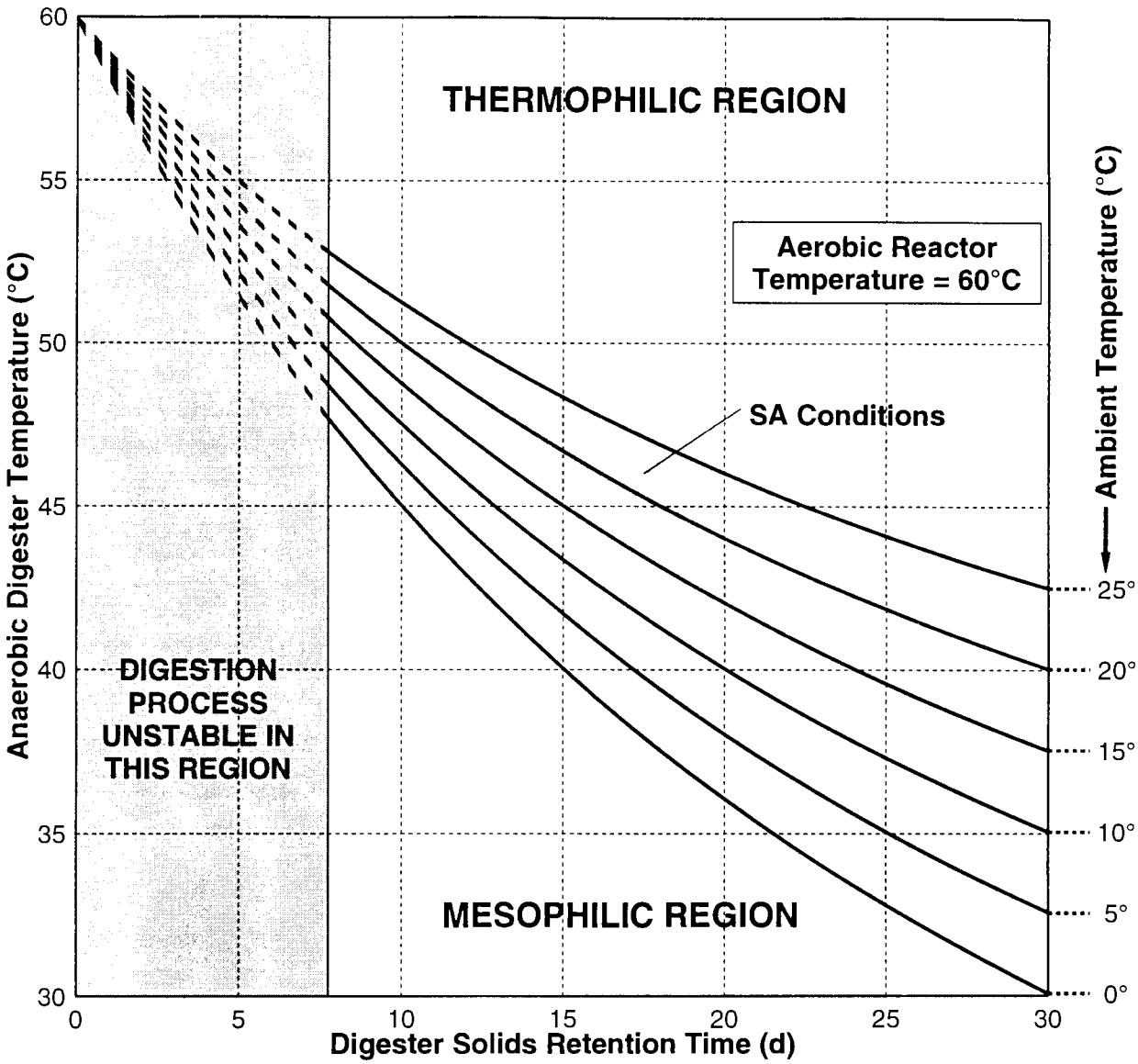


Figure 12a.2 The Variation in Forced Anaerobic Digester Operating Temperature with Digester Retention Time, for Different Ambient Temperatures; The Athlone Dual Digestion Plant, Aerobic Reactor Temperature Fixed at 60°C.

Figure 12a.2 illustrates clearly the danger and high probability of the anaerobic digester temperature falling between the mesophilic and thermophilic temperature regions when

operating a dual digestion plant. From an examination of Figure 12a.2, the following conclusion can be made with regard to the Athlone Dual Digestion Plant (this is also likely to be the case in general):

- For a digester retention time of 8-20 days with, (1) an ambient temperature of 15° to 25°C (South African temperatures) and (2) the aerobic reactor at 60°C (the minimum temperature for pasteurisation), **the digester temperature will be above the mesophilic region.**

From Figure 12a.2, it can be seen that there is a definite potential, with the relatively high ambient temperatures encountered in South Africa (~15° to 25°C) for the anaerobic digestion stage to be operated in the thermophilic region (50° to 60°). This aspect is explored further in the second application.

Application 2: Fixed Anaerobic Digester Temperature (53°C; Thermophilic)

In the second application, the anaerobic digester temperature is fixed at 53°C (the selected optimum temperature for thermophilic digestion). Rearranging Eq 12.4 in terms of reactor temperature, and substituting for $T(SL)_d$=53°, yields:

$$T(SL)_r = \frac{(0.008 \times 24 \times A_{dig} . R_h)}{(4.0 \times V_{reac})} . (53 - T_{amb}) + 53 \quad ...°C \quad (12a.8)$$

Substitution for the Athlone digester V_{dig}= 1800m^3 and A_{dig}= 1243m^2 into Eq 12a.8 yields:

$$T(SL)_r = \frac{(0.008 \times 24 \times 1243 . R_h)}{(4.0 \times 1800)} . (53 - T_{amb}) + 53 \quad ...°C \quad (12a.9)$$

which upon simplification yields:

$$T(SL)_r = 1.75 R_h - 0.0331 T_{amb} . R_h + 53 \quad ...°C \quad (12a.10)$$

Figure 12a.3 below shows the required aerobic reactor temperature $T(SL)_r$, to maintain the anaerobic digester temperature at 53°C for different digester retention times R_h and ambient temperatures T_{amb}. A similar application for an anaerobic digester temperature of 35°C is presented below, with the results also presented graphically in Figure 12.3.

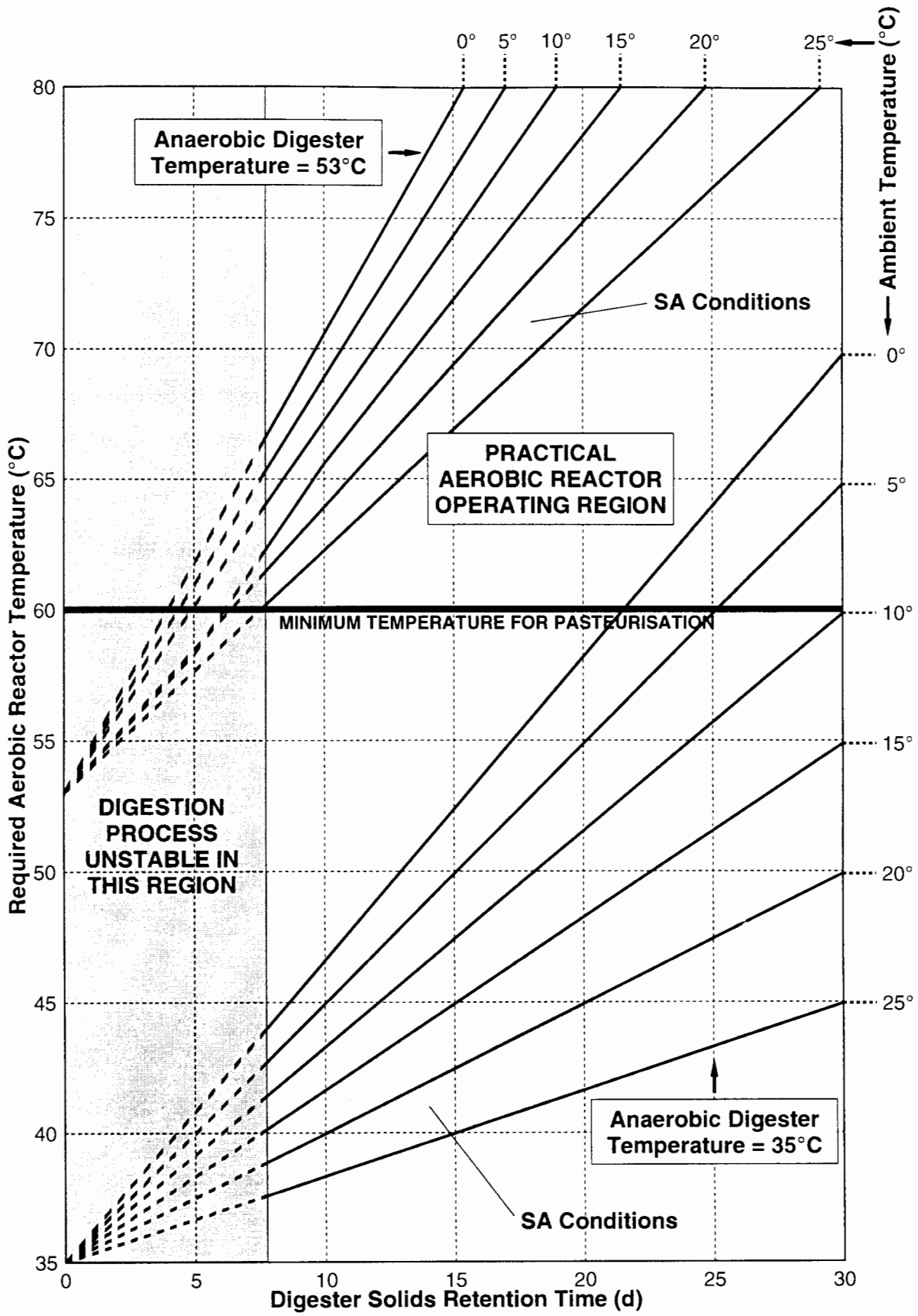


Figure 12a.3 The Variation in Required Aerobic Reactor Operating Temperature with Digester Retention Time, for Different Ambient Temperatures; The Athlone Dual Digestion Plant, Anaerobic Digester Temperature Fixed at 35°C (Mesophilic) and 53°C (Thermophilic).

Application 3: Fixed Anaerobic Digester Temperature (35°C; Mesophilic)

In the third application, the anaerobic digester temperature is fixed at 35°C (the selected optimum temperature for mesophilic digestion). Rearranging Eq 12.4 in terms of reactor temperature, and substituting for $T(SL)_d$=35°, yields:

$$T(SL)_r = \frac{(0.008 \times 24 \times A_{dig}.R_h)}{(4.0 \times V_{reac})}.(35 - T_{amb}) + 35 \quad ...°C \quad (12a.11)$$

Substitution for the Athlone digester V_{dig}= 1800m^3 and A_{dig}= 1243m^2 into Eq 12a.8 yields:

$$T(SL)_r = \frac{(0.008 \times 24 \times 1243.R_h)}{(4.0 \times 1800)}.(35 - T_{amb}) + 35 \quad ...°C \quad (12a.12)$$

which upon simplification yields:

$$T(SL)_r = 1.16R_h - 0.0331T_{amb}.R_h + 35 \quad ...°C \quad (12a.13)$$

From an examination of Figure 12a.3, it can be concluded that for the Athlone dual digestion plant (with an ambient temperature from 15° to 25°C), **the anaerobic digester temperature can be maintained in the thermophilic region (@53°C) at a retention times in the range 8 to 13 days.** The required aerobic reactor temperature will be in the range 60° to 70°C.

Conversely, trying to maintain the digester temperature at 35°C, without interstage heat exchange is not feasible for digester retention times below 20 days. The reactor temperature will be below that required for pasteurisation.

REFERENCES

Appleton AR and Venosa AD (1986) Technology evaluation of the dual digestion system. *J. Water Poliut. Control Fed.* 58 (16), 764-773.

Bruce AM and Oliver B (1987) Heating and cooling of sewage sludges - some recent developments. *Water Pollut. Control* 86 (1), 104-115.

Messenger JR, Ekama GA, de Villiers HA, Kenmuir K and Laubscher SA (1992) Evaluation and optimisation of dual digestion of sewage sludge - Part 2: Aerobic reactor performance. Final report WRC 189/3/92, Water Research Commission, PO Box 824, Pretoria, 0001.

APPENDIX 13

HEAVY METAL LEVELS IN THE ATHLONE FINAL SLUDGE

13a.1 Introduction

In terms of the recently published guidelines for the disposal of sewage sludge by land application in South Africa (DNH&PD, 1991), for a sludge to be classified as a Type D sludge (refer Section 1.2.1) and therefore suitable for unrestricted use, the heavy metal content in the final sludge must be below the levels listed in Table 13a.1 below.

Table 13a.1 Maximum Heavy Metal Content in mg/kg Dry Sludge for Classification as Type D in Terms of Guidelines DNH&PD (1991).

Heavy Metal	Limit (mg/kg)	Heavy Metal	Limit (mg/kg)
Cadmium	20	Cobalt	100
Chromium	1750	Copper	750
Manganese	-	Nickel	200
Lead	400	Zinc	2750
Molybdenum	25	Mercury	10

The primary sludge produced at Athlone is currently treated by conventional anaerobic digestion. In terms of the guidelines (DNH&PD, 1991) the treated sludge is classified as a Type B sludge and is therefore unsuitable for agricultural use. At present, this sludge is transported after drying to the Coastal Park Refuse Disposal Site. The negative aspects of this current practice are; (1) the beneficial use of the sludge as a soil conditioner and source of nutrients is lost, (2) the life span of the Refuse Disposal Site is reduced, and (3) significant costs are involved in transporting the sludge to the Disposal Site. The options available for the disposal of the sludge at Athlone would be greatly increased if that sludge were to be pasteurised. An increase in the monitoring of the influent raw wastewater at Athlone has placed tighter control on the discharge of abnormally high levels of heavy metals in Industrial Effluents. Heavy metal levels in the sludge are now below the maximum required levels stipulated by the guidelines (see Figure 13a.1). Accordingly if the sludge were to be pasteurised (i.e. by the dual digestion process) it

would classify as a Type D sludge and its disposal would not be restricted. Costs which are currently incurred in transporting the un-pasteurised sludge to the Refuse Disposal Site could be beneficially employed to provide pasteurisation. The production of a Type D sludge would prove attractive to those in agriculture who in all likelihood would provide transportation for the sludge.

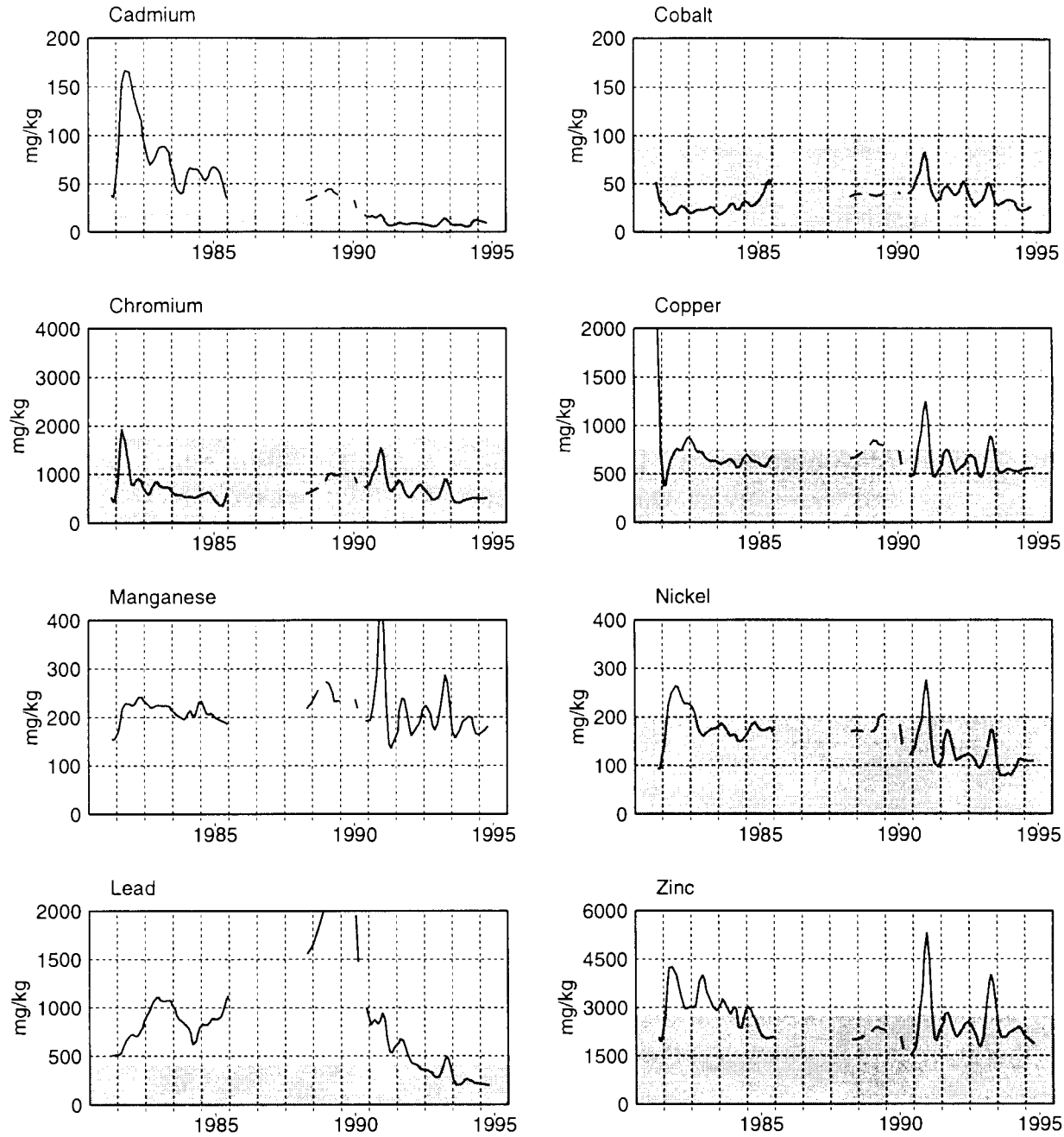
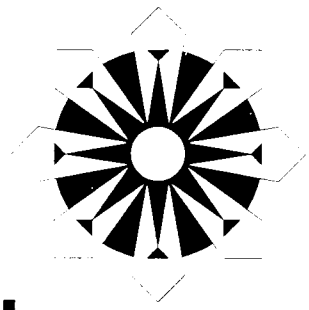


Figure 13a.1 Heavy Metal Levels in the Athlone Final Sludge: 1981-1995

REFERENCES

DNH&PD - Department of National Health and Population Development (1991) Guide: Permissible utilisation and disposal of sewage sludge. Ref. A11/2/5/4. Presented at 7th WISA Sludge Management Seminar, Aug., 1991.



APPENDIX 14

VACUUM SWING ADSORPTION- ON SITE OXYGEN PRODUCTION

14a.1 Oxygen Production from the Cryogenic Distillation of Liquid Air

Pure oxygen has traditionally been produced from the distillation of liquid air at cryogenic temperatures. The resulting purified oxygen is stored as a liquid at ambient temperature under pressure. The liquid oxygen is usually delivered by tanker, as and when required by the end-user, to be stored on-site in vacuum insulated pressure storage tanks. The liquid oxygen when required is passed from the tank through an evaporator. The reduction in pressure and the transfer of heat from the surrounding air, vaporise the liquid oxygen and bring it up to ambient temperature. A positive pressure is maintained for ease of application.

14a.2 Oxygen Production from Non Cryogenic Methods

Viable, non-cryogenic methods of oxygen production from the separation of air are now available. The two principle methods, which are of suitable for application in the dual digestion process, are *pressure swing adsorption* (PSA) and *vacuum swing adsorption* (VSA). These methods of oxygen production are an economic alterative to the liquid supply.

14a.3 Pressure Swing Adsorption

In the pressure swing adsorption process, a zeolitic molecular sieve which adsorbs nitrogen is used to separate the nitrogen from the other constituents in the air (principally oxygen).. The nitrogen builds up in the pore system of the zeolitic molecular sieve while the oxygen remains as a product gas. With the PSA process, oxygen purities of between 90% and 95% by vol. can be obtained. The PSA process is used for the production of oxygen quantities up to approximately 100 kg(O_2)/h, for which the specific energy required is approximately 1.2 kWh/kg(O_2).

14a.4 Vacuum Swing Adsorption

The vacuum swing adsorption process has become the preferred system for the production of large quantities of oxygen of up to 4000 kg(O_2)/h. Purities of between 90 and 99% can be obtained for which the specific energy required is approximately 0.5

kWh/kg(O_2). A schematic of a typical VSA plant is shown in Figure 14a.1. The principal components of the VSA plant are an air fan (compressor), air heater (for operation in winter), two or three adsorption vessels, vacuum pump and an oxygen compressor.

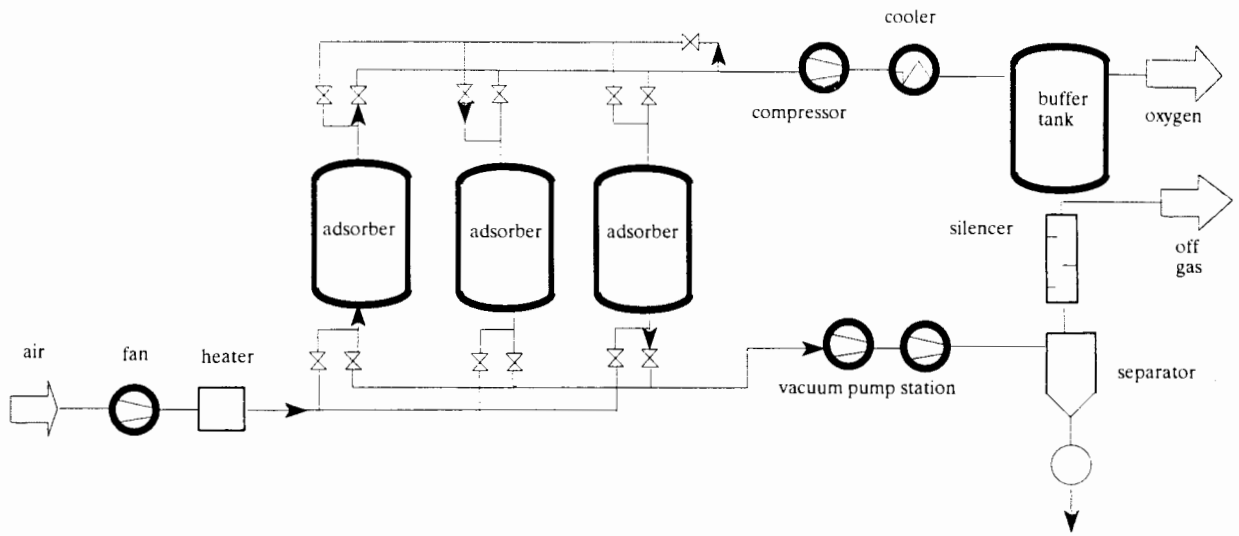


Figure 14a.1 Schematic Diagram of a VSA Plant for the Production of Oxygen

The separation of oxygen and nitrogen takes place at a slight over-pressure in the adsorption vessel (filled with a zeolitic molecular sieve material). This results in the enrichment of the air with oxygen when air flows through the vessel. To regenerate the molecular sieve, the vessel pressure is reduced with the help of a vacuum pump. For a continuous production of oxygen the adsorption vessels successively undergo pressure build-up (filling), adsorption and desorption (pressure reduction).

14a.5 Application to the Dual Digestion Process

The VSA process is ideally suited for application in the dual digestion process, which requires a continuous supply of oxygen that does not need to be high purity grade (>99.9%). Purities in the range 90-99% would be adequate. The VSA plant would be located on site. A liquid oxygen storage tank is usually installed alongside to provide a secondary source of oxygen in the event of mechanical failure or planned maintenance shutdown to the VSA plant or to provide an increased flow rate if and when required.

The procedure applied by the majority of the gas supply companies in South Africa is for the client to rent the VSA plant (maintenance of the plant remains the responsibility of the supplier). The current cost of purchasing oxygen in this manner is approximately half the liquid oxygen cost at around 0.25c/kg(O_2) (includes electricity costs; May 1995).

APPENDIX 15

OXYGEN SOLUBILITY

15a.1 Introduction

When injecting pure oxygen into the sludge recirculation line, it cannot simply be assumed that the pure oxygen injection equipment will be capable of delivering whatever oxygen transfer rate ***OTR*** is required for a specific set of operating conditions. In practice there will be a limit to the rate of oxygen transfer which can be effected. This limit is defined as OTR_{max}, the maximum oxygen transfer rate, and is principally dependent on the solubility of the pure oxygen at the point of injection. In order to predict OTR_{max} consideration must be given to the physical conditions (temperature and pressure) at the selected point of injection.

15a.2 Estimation of the Maximum Oxygen Transfer Rate

An estimate of the maximum oxygen transfer rate $M(O_2)^{max}_{trans}$ $(=V_p.OTR_{max})$ which can be effected, can be made from the product of the sludge flow rate and the solubility of oxygen at the temperature and pressure in the recirculation line viz.

$$M(O_2)^{max}_{trans} = C(O_2)^{sat}_{P,T} \times Q(SL)_{mix} \quad \text{...kg}(O_2)/\text{h} \quad (15a.1)$$

where:

$M(O_2)^{max}_{trans}$ = Maximum oxygen transfer rate (kg(O_2)/h)
$C(O_2)^{sat}_{P,T}$ = Oxygen solubility at pressure P and temperature T (kg(O_2)/m^3)
$Q(SL)_{mix}$ = Sludge flow rate in recirculation line (m^3/h)

The variation in solubility of oxygen in pure water with temperature under a total pressure of 760 mmHg is shown in Figure 15a.1 below (data obtained from the Handbook of Chemistry and Physics, 13th Ed., 1948). The total pressure is equal to the partial pressure of the oxygen plus the aqueous surface tension.

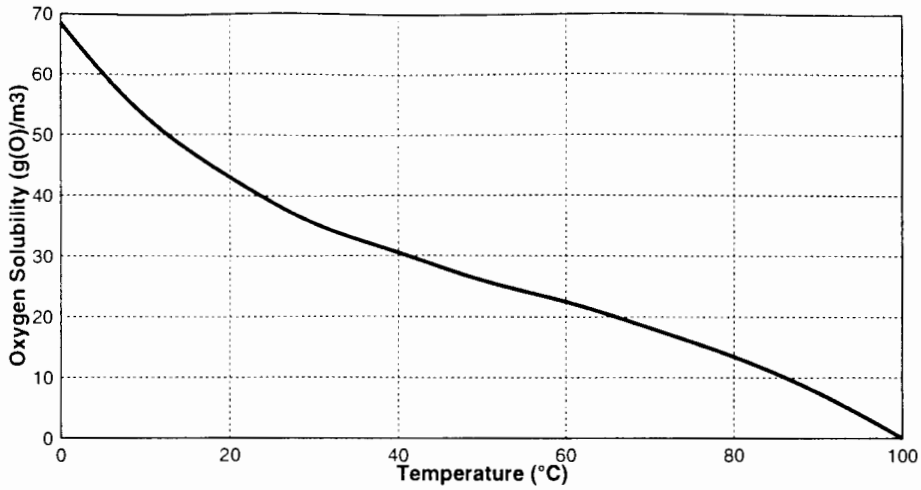


Figure 15a.1 The Variation in the Saturation Concentration (Solubility) of Oxygen in Water with Temperature under a Total Pressure of 760mmHg.

The temperature of the sludge in the recirculation line will be in the range 50° to 65°C. From Figure 15a.1, at an oxygen partial pressure of 760mmHg, the oxygen solubility at 50°C and 60°C is as given in Table 15a.1 below.

Table 15a.1 The Solubility of Oxygen in Pure Water at 50°C and 60°C ($kg(O_2)/m^3$) Under a Total Pressure of Oxygen of 760mmHg (Including Surface Tension).

Temperature (°C)	Solubility ($kg(O_2)/m^3$)
50°	0.0266
60°	0.0227

If the oxygen partial pressure is increased, then the solubility of the oxygen in the water will increase in proportion according to the following relationship:

$$C(O_2)^{sat}_{P,T} = \frac{P(O_2)}{760} \times C(O_2)^{sat}_{760,T} \quad \text{...kg}(O_2)/m^3 \quad (15a.2)$$

where:

$P(O_2)$ = Oxygen partial pressure (mmHg)

$C(O_2)^{sat}_{760,T}$ = Oxygen solubility at pressure 760mmHg and temperature T ($kg(O_2)/m^3$)

Consequently, the maximum oxygen transfer rate $M(O_2)^{max}_{trans}$ in terms of sludge flow rate and pressure can be obtained by substituting Eq 15.2 into Eq 15.1 viz:

$$M(O_2)^{max}_{trans} = \frac{P(O_2)}{760} \times C(O_2)^{sat}_{760,T} \times Q(SL)_{mix} \quad \text{...kg}(O_2)/h \quad (15a.3)$$

To generate the necessary pressure and flow rate in the sludge recirculation line to induce a sufficiently high maximum oxygen transfer rate, the design of the pipework and the sizing of the pump(s) requires careful consideration. For Athlone the design and subsequent modification made to the recirculation line, to allow for pure oxygen injection during phase II, is described in Appendix 10. During the phase II investigation period, the flow characteristics required to achieve a pure oxygen transfer efficiency *OTE* of 80% were determined (refer Section 7.3.3). These flow characteristics are contained in Table 15a.2 below.

Table 15a.2 Minimum Required Flow Characteristics in the Sludge Recirculation Line to Achieve a Pure Oxygen Transfer Efficiency OTE^{O2} of 80%

Minimum Required Recirculation Sludge Flow Rate (m^3/h)	Velocity at the Point of Discharge (m/s)	Pressure Head Generated in the Recirculation Line (m)	Minimum Required Pure Oxygen Transfer Efficiency (%)
690	13.2	23.0	**80.0**

The total pressure in the recirculation line at the point of injection, and therefore the pressure at which pure oxygen is injected into the sludge stream, is given by:

$$P(O_2) = (h_s + h_f).\frac{760}{10.3} \quad \text{...mmHg} \quad (15a.3)$$

where:

h_f = Pressure head generated in the recirculation line (= 23.0m)

h_s = Static head (atmospheric + hydrostatic ≡ 10.3 + 6.0 = 16.3m)

760/10.3 = Conversion factor (m→ mmHg)

Substitution of the values for h_s and h_f into Eq 15a.3 yields:

$$P(O_2) = (16.3 + 23.0).\frac{760}{10.3} = 2900 \quad \text{...mmHg} \quad (15a.4)$$

Consequently, at a sludge flow rate $Q(SL)_{mix}$ of 690m^3/h and oxygen pressure $P(O_2)$ of 2900mmHg, the calculated maximum oxygen transfer rate $M(O_2)^{max}_{trans}$ (from Eq 15a.3) for a sludge temperature of 60°C (where $C(O_2)^{sat}_{760,60°}$ = 0.0227 kg(O_2)/m^3) is given by:

$$M(O_2)^{max}_{trans} = \frac{2900}{760} \times 0.0227 \times 690 = 59.8 \quad \text{...kg}(O_2)/\text{h} \quad (15a.4)$$

This gives a value for the volume specific maximum oxygen transfer rate OTR_{max} of (59.8/184) **0.325kg(O_2)/m^3.h**.

15a.3 Comparison with Actual Operating Data

The highest pure oxygen transfer rate ***OTR*** recorded during phase II occurred during steady state period 10, with a measured ***OTR*** = **0.435 kg(O_2)/m^3.h** (refer Section 7.3.2). At this time the sludge flow rate $Q(SL)_{mix}$ was estimated at 700m^3/h with an oxygen pressure $P(O_2)$ of 2930mmHg, the calculated maximum oxygen transfer rate $M(O_2)^{max}_{trans}$ from Eq 15a.3 = 61.3 kg(O_2)/h, giving an OTR_{max} of (61.3/184) **0.333 kg(O_2)/m^3.h**.

The highest recorded ***OTR*** (0.434 kg(O_2)/m^3.h) is 30% greater than the estimated value for OTR_{max} (0.333 kg(O_2)/m^3.h)[1]. This could be attributable to either;

(1) supersaturation taking place in the sludge recirculation line, or
(2) the shattering effect on the undissolved oxygen bubbles, when the sludge is discharged back into the reactor at high velocity (~13.4m/s), causing this oxygen to be subsequently dissolved (as the already dissolved oxygen is utilised by the bacteria).

Note that, it is accepted that a significant proportion of the oxygen, which is dissolved into the sludge liquid under pressure in the recirculation line, will come out of solution as the pressure is released upon re-entry into the reactor. Indeed, one of the built in design features of the sludge recirculation line is the necessity to have a high discharge velocity to produce the shattering effect on the gaseous oxygen as it comes out of solution. A fine 'mist' of gaseous oxygen is produced which is readily re-dissolved. The higher ***OTR*** may well be a consequence of the merger (and shattering together) of un-dissolved oxygen and the 'dissolved' oxygen which comes out of solution upon re-entry.

REFERENCES

Handbook of Chemistry and Physics, 13th Ed., (1948) Edited by CD Hodgman, Chemical Rubber Publishing Co.

[1] At this ***OTR*** (=0.435 kg(O_2)/m^3.h) the aerobic reactor was still oxygen limited, indicating that higher ***OTR***'s could have been possible up to the limit OTR_{max} .